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ADVANCED WASTE TREATMENT SEMINAR, SESSION III, REMOVAL OF SOLIDS AND ORGANICS, HELD AT SAN FRANCISCO, ON OCTOBER 28-29, 1970

Federal Water Quality Administration

Prepared for:

California State Water Resources Control Board

October 1970

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CLEAN WATER

SESSION III

REMOVAL of SOLIDS & ORGANICS

ADVANCED WASTE TREATMENT SEMINAR

San Francisco
October 28 & 29
1970

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KABUKI THEATRE
1881 POST STREET
SAN FRANCISCO

OCTOBER 28-29, 1970

JOINTLY SPONSORED BY
CALIFORNIA STATE WATER RESOURCES CONTROL BOARD
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PROGRAM

OCTOBER 29 MORNING

SESSION III REMOVAL OF SOLIDS AND ORGANICS

Moderator: Professor Ray B. Krone

9:00 AM Solids Removal Processes - Jesse M. Cohen

9:30 AM New Developments in Sludge Handling - Robert B. Dean

10:15 AM Removal of Soluble Organics on Carbon - John N. English

10:45 AM Use of Pure Oxygen in Biological Treatment
Edwin F. Barth

11:15 AM Summary - Professor Krone

CONTENT

Solids Removal Processes.....Jesse M. Cohen

New Developments in Sludge Handling and Disposal.....Robert B. Dean

Removal of Organics from Wastewater by Activated Carbon.....
English, Masse, Carry, Pitkin and Haskins

Investigation of the Use of High Purity Oxygen Aeration in the Conventional Activated Sludge Process (Summary).....Edwin F. Barth

The Seminar also included the below sessions:

Session I	Nitrogen Removal
Session II	Phosphorus Removal
Session IV	Combined Treatment and Application

SOLIDS REMOVAL PROCESSES

Jesse M. Cohen

If one were to rank wastewater treatment operations in order of overall importance, separation of solids from liquids would probably assume major importance. This operation may be required to remove objectionable components from a waste, to render the waste more amenable to subsequent treatment or disposal, or to support another treatment operation.

SEDIMENTATION

Sedimentation tanks are employed for various applications which have one common objective - the removal of solid matter from a flowing liquid. The solid matter may have been present in direct suspension, such as in municipal sewage, or may alternatively be a precipitate resulting from prior chemical treatment, such as lime precipitation or alum or iron coagulation.

Horizontal Flow Design

The principal basis of the earliest designs of horizontal-flow sedimentation tank was the "retention time". The objective was to move water so slowly through the tank that there would be ample time for settlement. The nominal detention period allowed was usually more than four hours. However, each drop of water did not take the same time to travel from inlet to outlet, since some degree of "short-circuiting" was bound to occur. In fact, a typical pattern of flow under ideal conditions in an elementary design of a horizontal-flow tank is represented by Fig. 1. Diagram (a), solid line, represents the most favored flow-path, while the dotted lines indicate typical eddies which are induced in the remainder of the tank volume. Diagrams (b) and (c) represent the flows which result, respectively, to entering water being warmer or colder than the water in the tank.

The basic design which has been most widely used for sewage treatment is a tank circular in plan. The inflow is introduced at the center of the tank and the outflow is collected at a peripheral launder. As the general direction of flow is mainly a radial spread from the center, it is reasonable to regard this design as in the horizontal-flow category.

Presented at the Advanced Waste Treatment Seminar, Pacific Southwest Region, FWQA, San Francisco, California, October 28-29, 1970.

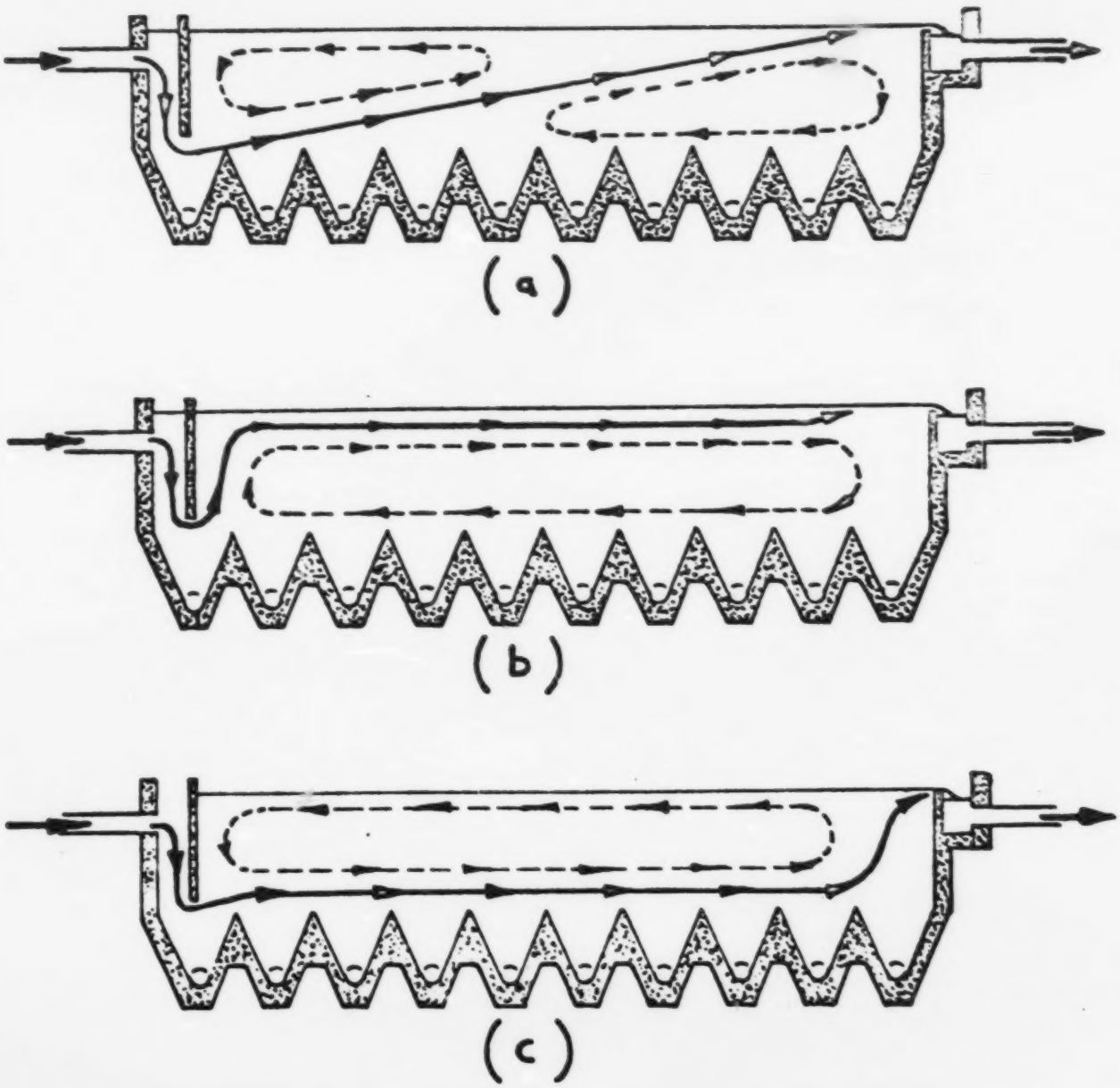


Figure 1—Sedimentation tank of horizontal-flow type under different conditions of relative temperature.

Vertical-Upflow Design

The improved performance obtained in upflow tanks has led to a variety of new designs, incorporating variants of the upflow principle. These new designs are designated by various proprietary names - Accelator, Clarifow, Flocsettler, Reactor-Clarifier, etc. One of the earliest of the upflow tanks was the Spaulding Precipitator, shown in Fig. 2. In this design, the flow is introduced into the center of the tank and flows upward through a blanket of previously formed solids.

The principal advantages of the upflow versus the horizontal tanks are (1) improved flow control, and (2) sludge blanket effect. Salt-injection tests performed on a variety of tank designs clearly show the superior efficiency of the upflow principle in the matter of flow control. The radial-flow still largely used in sewage treatment is distinctly inferior, being subject to short-circuiting.

Tube and Lamella Settlers

In an ideal settling basin, as defined by Camp, the paths of all discrete particles will be straight lines, and all particles with the same settling velocity will move in parallel paths. The settling pattern shown in Fig. 3 would be the same for all longitudinal sections. It is apparent from this that as the interval (h) is reduced, the size of the basin required to remove a given percentage of the incoming settleable material decreases. Many devices have been proposed since this principle was first proposed by Hazen in 1904 and further developed by Camp in 1946. None of the proposed devices were accepted commercially until the recent introduction of the device called the tube settler. These consist essentially of closely packed small diameter tubes, 1-4 inches in diameter and 2-4 feet in length, inclined at some angle to provide for removal of sludge as the water flows upward through the tubes. Detention times are in the order of 6 minutes and less. The tubes provide as much as 24 hours of sludge storage, depending of course, on the amount of suspended solids, and sludge is readily removed by gravity drainage. A schematic diagram of the tube settler integrated into a complete clarification device which consists of coagulation-flocculation, a tube settler unit and a mixed media filter is shown in Fig. 4. More recently, a device consisting of parallel plates rather than tubes has become available. The device is called a lamella. Flow is co-current in contrast to the countercurrent flow of the tube settlers. The plate arrangement in a tank is shown in Fig. 5.

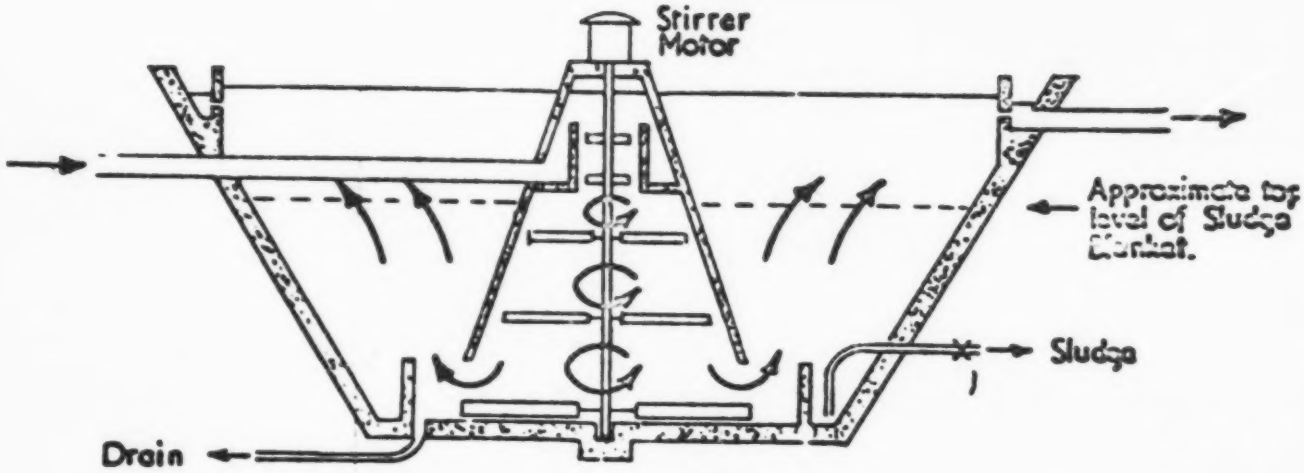


FIG. 2 — *Spaulding Precipitator*

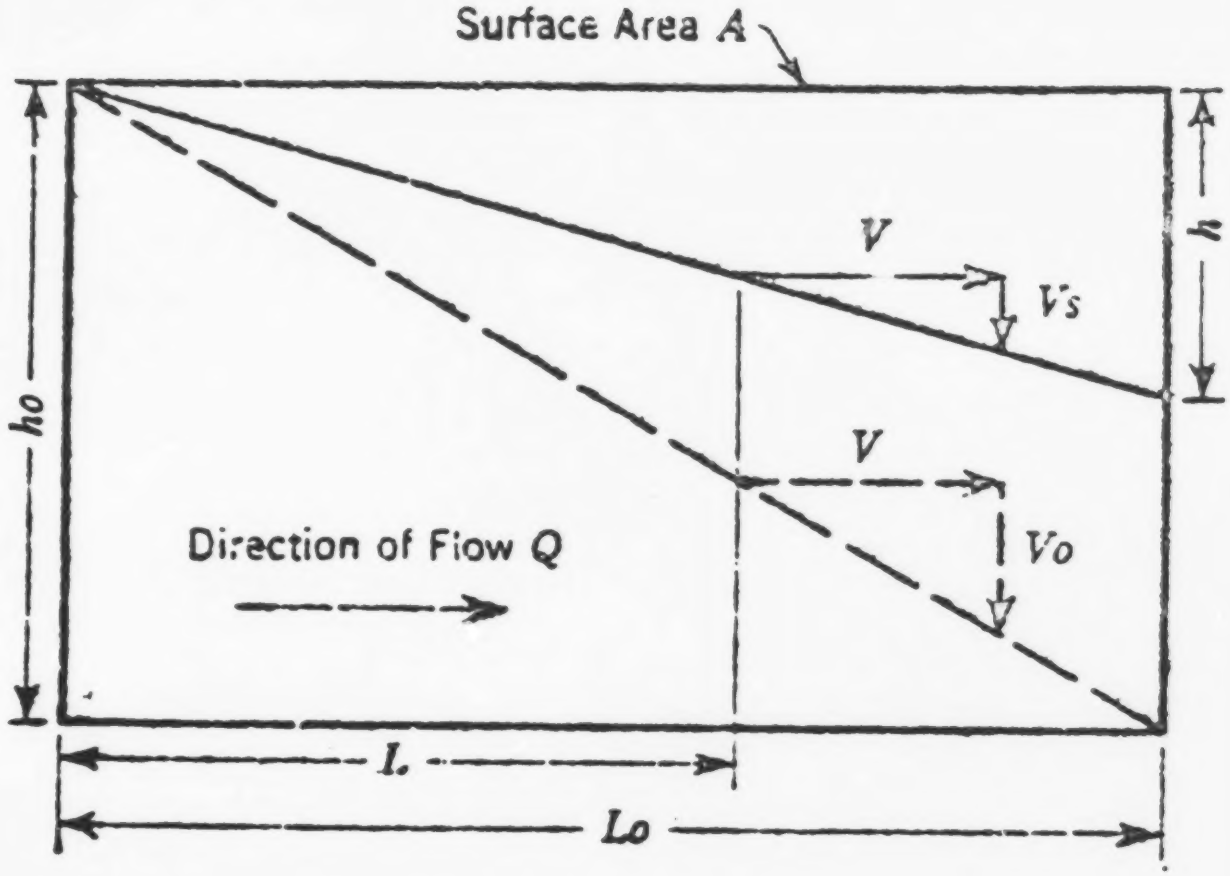


FIG. 3 **Idealized Settling Paths of Discrete Particles in a Horizontal Flow Tank**

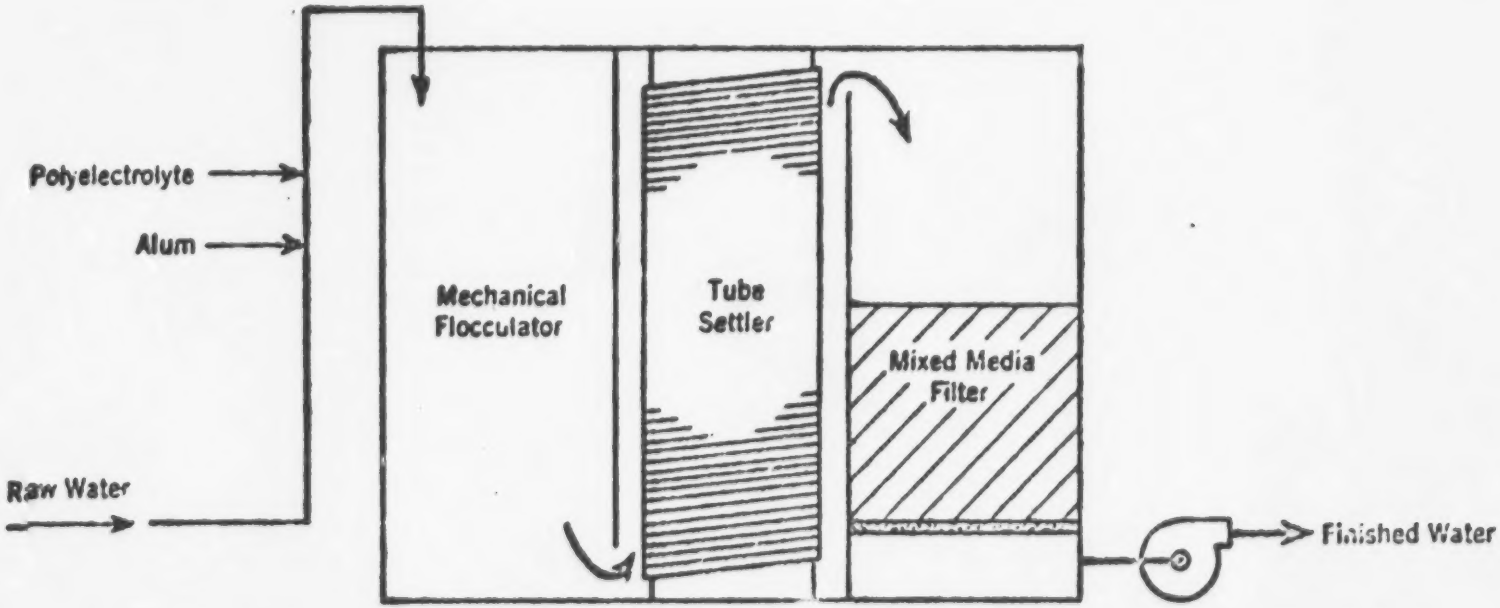
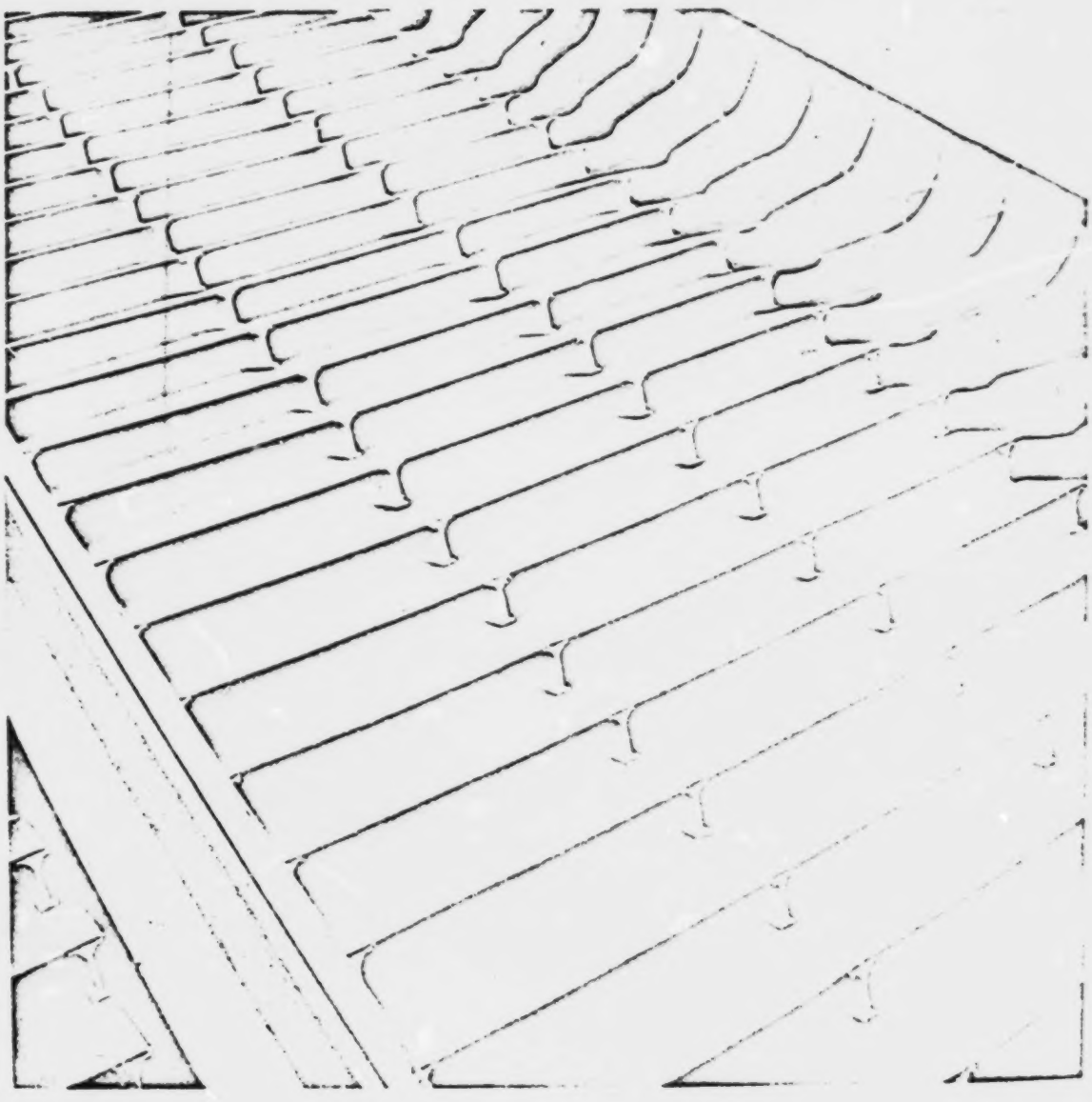


Fig. 4. Schematic Diagram of Apparatus Used in Field Tests of Tube Settler

Figure 5 Lamella Separator



DISSOLVED AIR FLOTATION

The use of dissolved air to float suspended solids was first used in industrial operations. In recent years, the process has been adapted to domestic wastewater and particularly to sludge thickening. The process achieves the separation of suspended particles by attachment of gas bubbles to the suspended particles, thereby reducing the effective specific gravity of the particles to less than that of the water.

A. A flotation system using the pressurization and de-pressurization sequence consists of the following elements (Fig. 6).

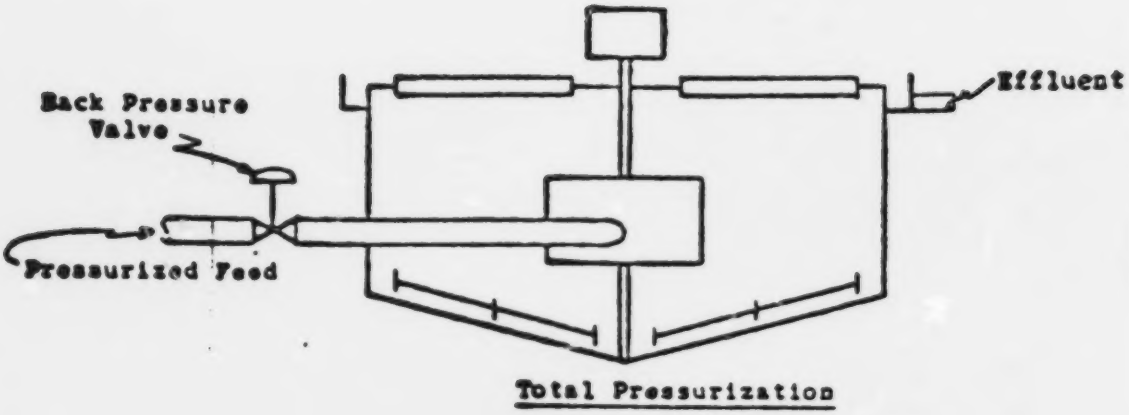
1. Pressurizing pump
2. Air injection facilities
3. Retention tank or contact vessel
4. Back-pressure regulating device
5. Flotation device
6. Facility for addition of chemicals if needed

B. Advantages and Disadvantages

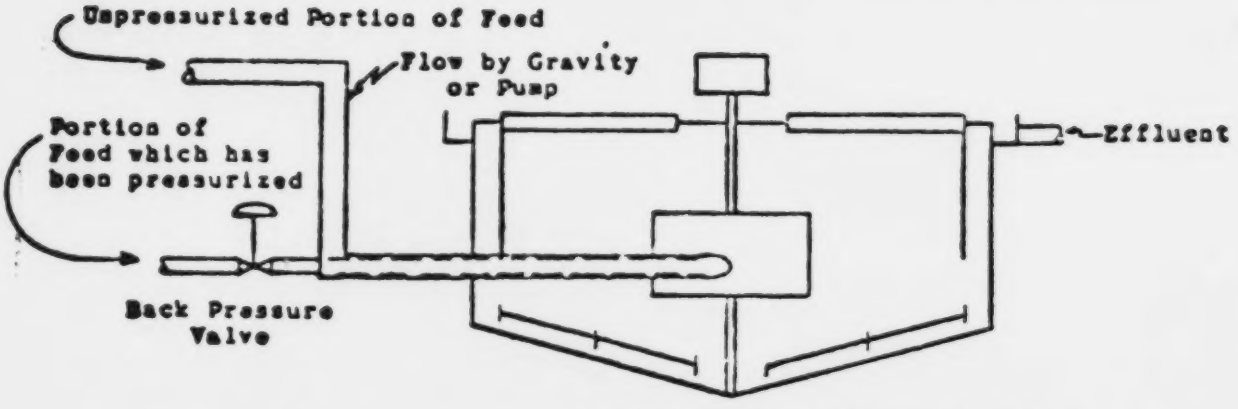
1. Much reduced retention time - 10-20 minutes
2. Greater solids concentration in float than in settled sludge
3. Greater efficiency of solids recovery
4. Offers mechanical control over the process
5. Increased cost of operation for pumping, etc.
6. Need to remove top and bottom solids

SCREENING DEVICES

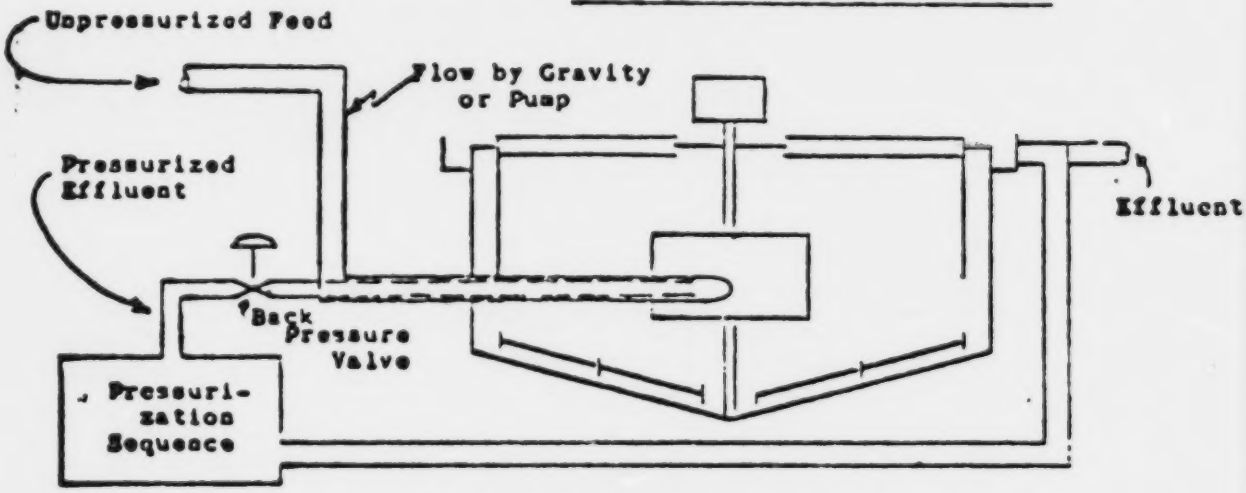
Microscreening is a form of simple filtration by straining (Fig. 7). These mechanical filters consist of a rotary drum which revolves on a horizontal axis. The peripheral surface of the drum is covered with a stainless steel fabric. The effectiveness of the woven mesh screen for retaining fine particles is dependent on the size of the openings in the screen and on the pattern of the weave. Influent enters the open end of the drum and is filtered through the fabric with the intercepted solids being retained on the inside surface of the fabric. As the drum rotates, the solids are transported and continuously removed at the top of the drum by pumping strained effluent, under pressure, through a series of spray nozzles which extend the length of the drum. The solids and wash water are collected in a central trough within the drum and discharged through a hollow axle. The microstraining device is available in several unit sizes ranging from 5' in diameter and 1' width with a capacity of 0.05 to 0.5 mgd to 10' diameter and 10' wide with a capacity of 3-10 mgd.



Total Pressurization


Unpressurized Portion of Feed
Flow by Gravity or Pump
Portion of Feed which has been pressurized
Effluent
Back Pressure Valve

Partial Pressurization of Feed



Partial Pressurization of Effluent

Methods employed for partial and total pressurizations.

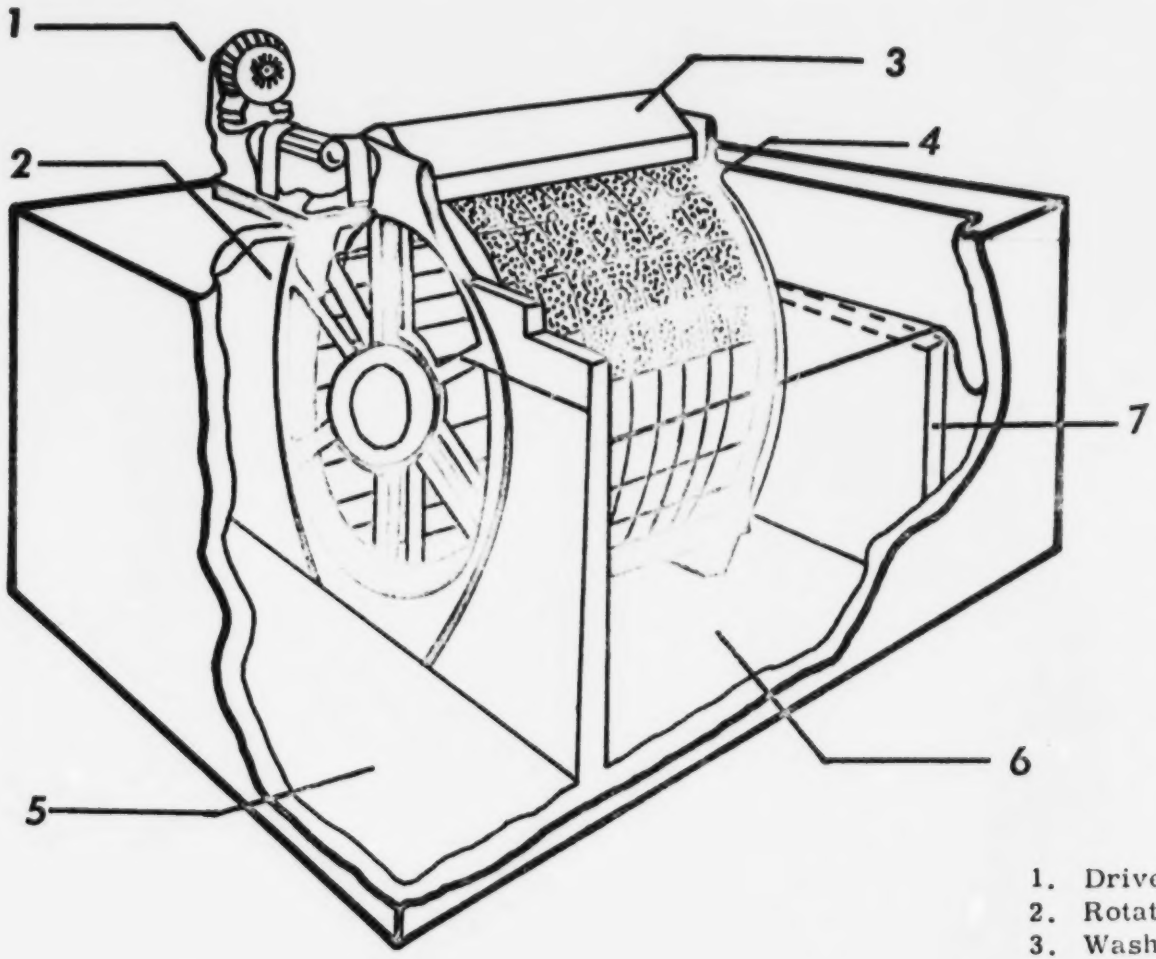


Fig. 1 Typical Microstrainer Unit

1. Drive Unit
2. Rotating Drum
3. Wash-water Jets
4. Micro-fabric
5. Influent Chamber
6. Effluent Chamber
7. Effluent Weir

Not Shown

Wastewater Hopper
Ultra-violet Lamp
Wash-water Pump

FIGURE 7

Microscreening devices have found their greatest application to treatment of river waters, and information on their performance on wastewaters is quite scarce. The effect of aperture size on removal efficiencies and flow rate is shown below.

Fabric	Removal Efficiency		Flow
	Solids	BOD	gals/hr/ft^2
Mark 0 (23 microns)	70-80%	60-70%	400
Mark 1 (35 microns)	50-60%	40-50%	600

The advantages of microstraining are the low initial capital cost and ease of operation. The disadvantages are the incomplete solids removal and the inability to handle solids fluctuations.

Capital and Operating Costs

The capital and operating costs associated with microstraining have been prepared by Smith and are shown in Fig. 8. For a 10 mgd plant, total cost for microstraining is calculated to cost about 1.5¢/1000 gallon.

In-Depth Filtration

In-depth filtration is the passage of a fluid through a bed of granular media designed to permit the captured particles to be retained within the filter. The degree of penetration and solids removal efficiency can be altered by changing the size and character of the particulates to be removed as well as the size and composition of the filter media itself.

A basic prerequisite for the operation of rapid sand filters is that good coagulation and flocculation must be obtained. Barring adequate pretreatment of the wastewater, filtration efficiency is decreased as evidenced by "breakthrough" of floc. With good pretreatment by coagulation, higher filtration rates are attainable while still maintaining clarity of the effluent.

It is beyond the scope of this talk to discuss the subject of coagulation and flocculation. A variety of coagulants and flocculants are available including various salts of aluminum and iron, lime and organic polymeric flocculants. The former inorganic salts, in the proper dosages can also provide for precipitation of phosphate in addition to clarification.

MICROSTRAINING OF SECONDARY EFFLUENT

Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity

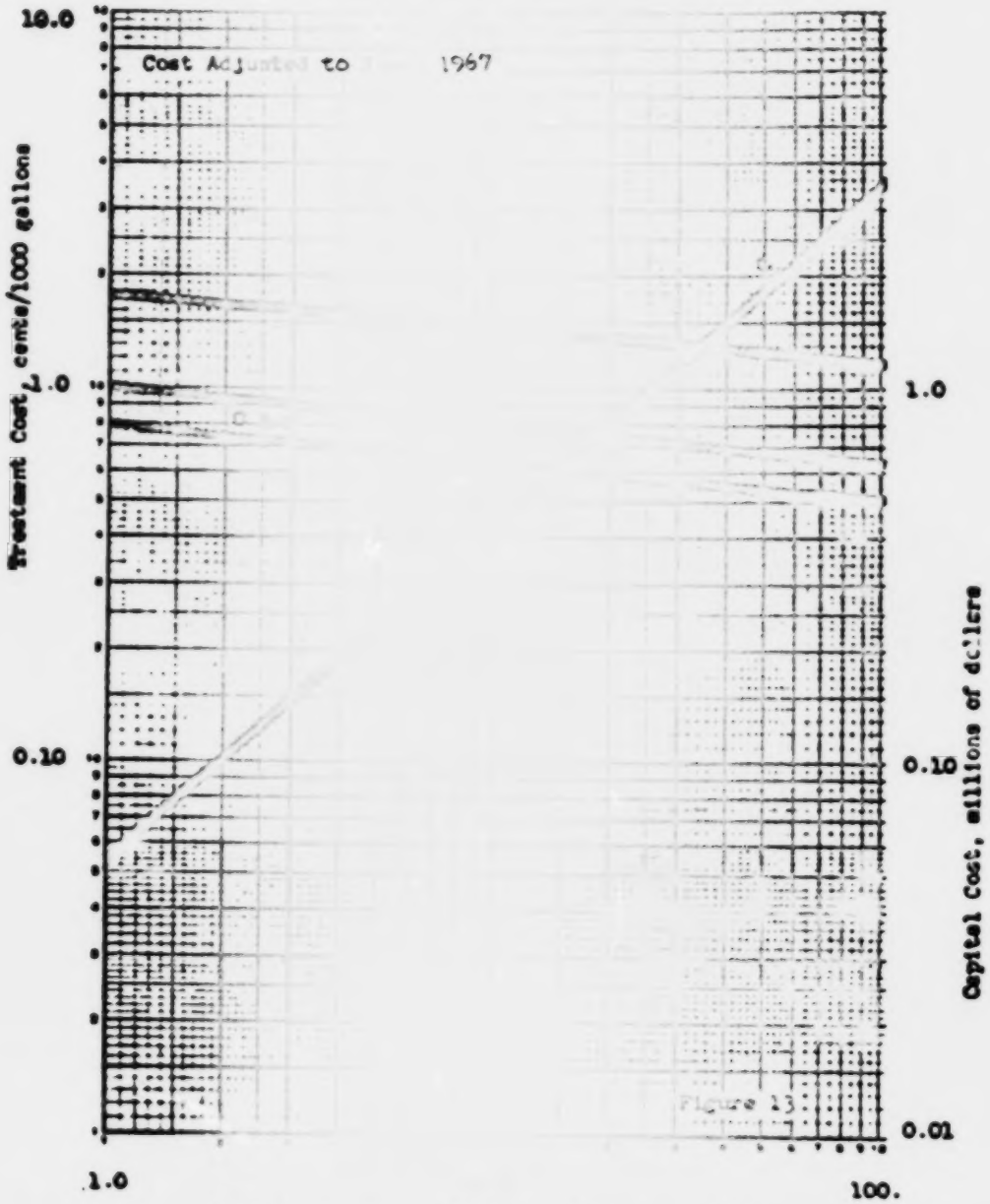


Figure 13

Reproduced from best available copy.

C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons (4 1/2% - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

FIG. 8

In the past 70 or 80 years there has been a gradual improvement in the basic process of media filtration. Some of the earlier work centered on increasing the filtration rates of slow sand filters to the modern rapid sand filter rates. Various cleaning methods were also developed including backwashing and surface scouring. More recently, the engineering advances have been concerned with modifications of filter media that would allow greater production of high quality water from a given filter area.

The evolution of filter design is illustrated in Fig. 9. The cross-section shown at the top represents the rapid-sand filter which is in common use in many filter plants today. Typical effective size of the media used is 0.5 mm although effective sizes from 0.35 to 1.0 mm have been used. During filter backwashing, the sand grades hydraulically with the finest particles rising to the top of the bed. As a result, most of the material removed by the filter is removed at very near the surface of the bed. Only a small part of the total voids in the bed are used to store particulates and headloss increases very rapidly. When secondary effluent is being processed, the high solids concentration will blind the surface in a very short time. As much as 75-95% of the headloss, under these conditions, will occur at the upper 1-inch layer of the filter. Filter runs will be so short as to be prohibitive. Further, floc breaking through the topmost layers, have increased opportunity to pass through the entire filter since voids become increasingly larger with increase in depth.

One approach to increasing the effective filter depth is the use of a dual media bed using a discrete layer of coarse coal above a layer of fine sand. The filter provides basically a two-layer effect to achieve increased penetration of particles. The amount of sand is reduced to afford lower headlosses at the higher throughput rates used. Normally, such a bed is designed so that 24 inches of anthracite coal, with a nominal size of about 1 mm, overlays a 6 inch sand layer with a size of about 0.45 mm. Hydraulic stratification still occurs following backwashing but the difference of specific gravity is such that the larger coal remains on top of the sand. The bulk of filtration is accomplished in the upper layers of the coal and at the top inch or two of the sand bed.

With low applied turbidities and constant rate operation, the coal-sand media bed has demonstrated an ability to operate in the range of 4-5 gal/min/sq/ft of filter surface area. A defect in this design is that if a flow change occurs, the particles held in the relatively large void volume of a coal bed, can become dislodged and will be captured by the fine layer of sand and the filtration run would have to be terminated either because of high head loss at the coal-sand interface or because of breakthrough of particles through the relatively shallow sand bed. This design, then, presents a serious inconsistency in design.

Ideally, the effluent should pass through as fine a filter material as is feasible. This ideal design is illustrated in the bottom cross-section of a filter uniformly graded from coarse to fine from top to bottom.

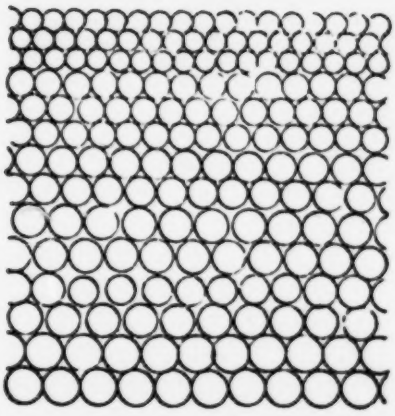
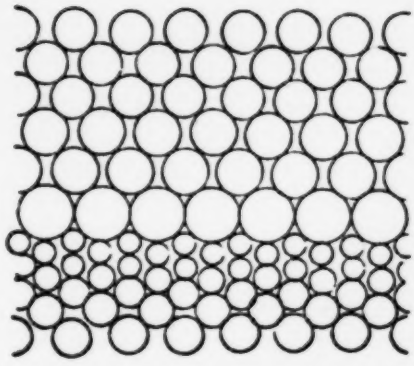


Figure 1
Cross-Section Through
Single-Media Bed
Such as Conventional
Rapid Sand Filter



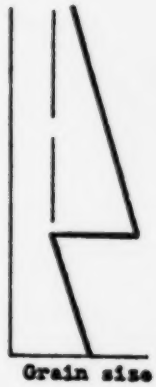


Figure 2
Cross-Section Through
Dual-Media Bed
Coarse Coal Above
Fine Sand

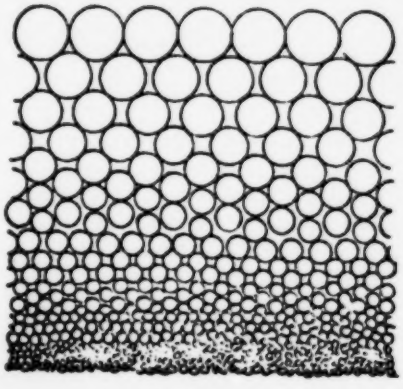


Figure 3
Cross-Section Through
Ideal Filter
Uniformly Graded From
Coarse to Fine
From Top to Bottom

FIG. 9

One solution is the use of three materials of differing specific gravity of such size gradation that some intermixing of the materials occur at the interfaces of the bed layers. The third media material is garnet which has a specific gravity of about 4.2. An ideal filter bed, then, would consist of about 60% of anthracite with a size of 1.0 to 1.5 mm at the top, 30% sand with a size of 0.4 to 0.5 mm at the middle and 10% of garnet with a size of about 0.15 mm. The materials are so sized that intermixing occurs at the interfaces. In this ideal filter, the effluent is passed through increasingly finer media. The uniform decrease in media particle size with filter depth allows the entire filter depth to be used for floc removal and storage.

Cost of the filtration step is shown in Fig. 10. For a 10 mgd plant this cost amounts to about 3.5¢/1000 gallon, when operating the filter at 4 gpm/sq ft. Further economy, of course, can be obtained at higher rates. Rates as high as 6-8 gpm/sq ft have been shown to be feasible when proper pretreatment of coagulation-flocculation followed by sedimentation is practiced.

Moving Bed Filter Technique

A new filtering technique has been evaluated by Johns-Manville under contract with FWQA. The technique is designed to overcome the problem of surface clogging and to achieve what is obtainable by multi-media filtration. A schematic diagram of the moving bed filter is shown in Fig. 11.

The unit is basically a sand filter. Particulate matter is removed as the water passes through the sand (0.6 to 0.8 mm). As the filter surface becomes clogged, the filter media is moved forward by means of a mechanical diaphragm. The clogged filter surface is removed either hydraulically or, as shown, mechanically thereby exposing a clean filter surface. The sand and accumulated sludge is collected and washed. The sand is returned via a hopper to the base of the bed. The unit is thus a form of countercurrent extraction device feeding sand countercurrent to the water being filtered. The moving bed filter has a renewable filter surface analogous to the microstrainer and the advantage of depth filtration comparable to the coarse media filter. The unit does not have to be taken off-stream for backwashing. In theory, 1% of the filter is being backwashed 100% of the time compared to the conventional practice of backwashing 100% of the filter 1% of the time.

Several pilot MBF units have been built to date and used to treat settled and non-settled trickling filter effluents and primary effluent. The system lends itself well to the use of chemical aids ahead of filtration because of designed flexibility to handle high solids loadings.

FILTRATION THROUGH SAND OR GRADED MEDIA - 4GPM/SQ FT

Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity

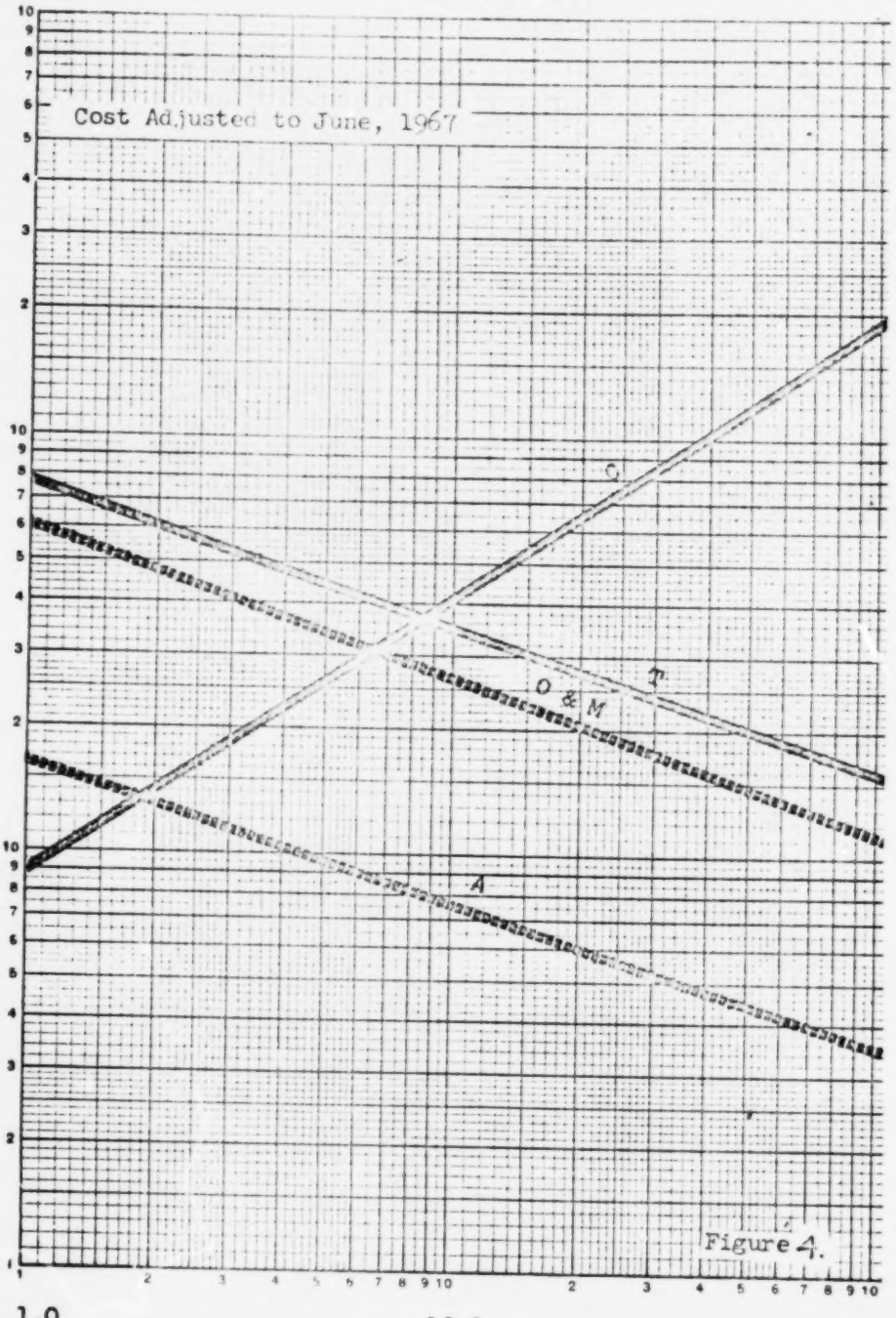


Figure 4.

C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons(4 1/2% - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

BASIC CONCEPT OF MOVING BED FILTER

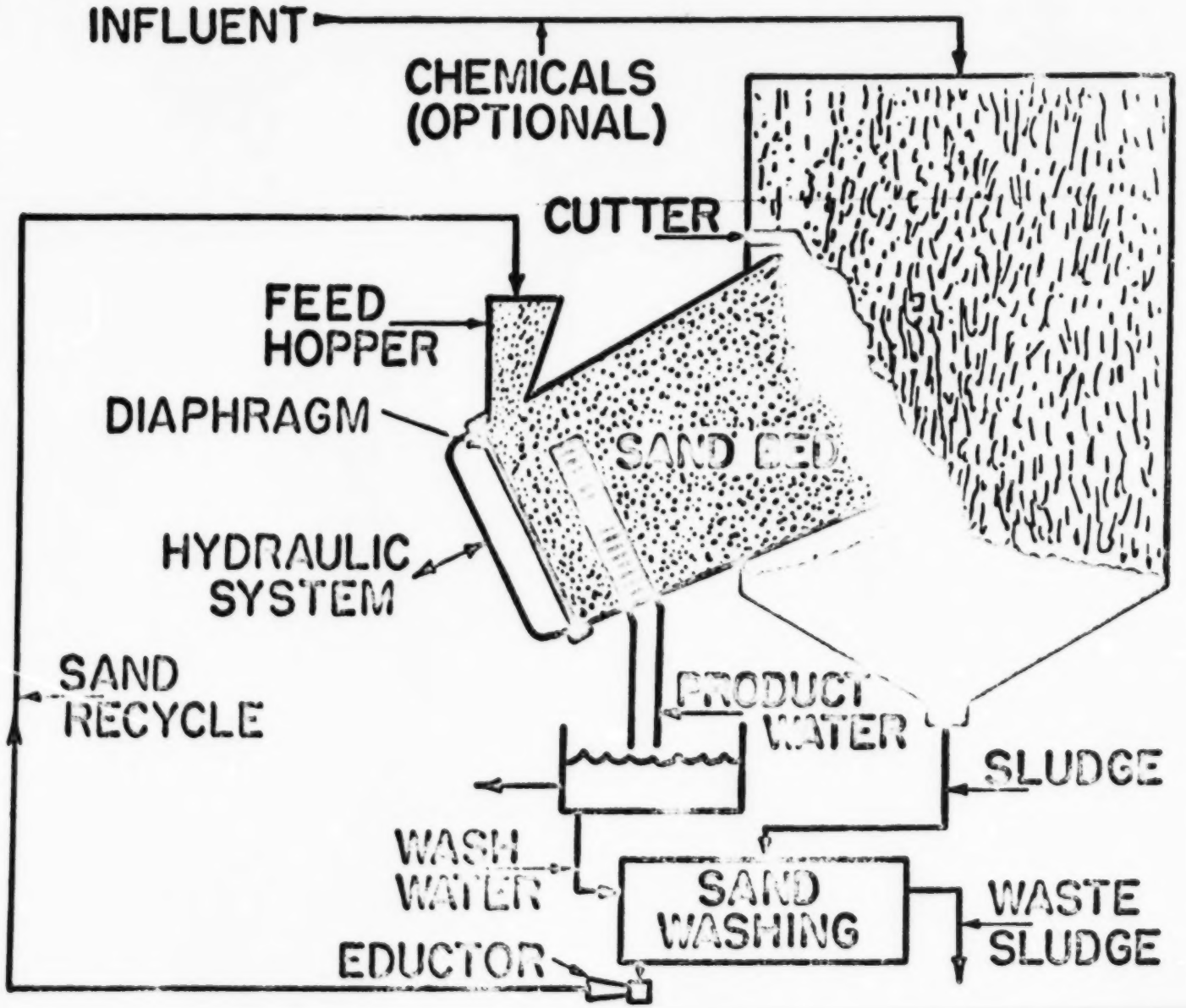


Fig. 12 shows some data with the MBF treating non-settled trickling filter effluent with various dosages of alum and coagulant aid. Product quality was maintained at about 10 BOD while being fed an influent with a widely varying BOD content of ≈40 to as high as 180 BOD.

It is too premature to talk about cost of this method of filtration. Design and performance information developed from these pilot units will be used to obtain these costs.

ULTRAFILTRATION

One of the newest unit processes to separate solids from liquids is the operation known as ultrafiltration. While this method has been under development over the past 10 years, it is only within the past two-three years that has seen the commercial development of this process.

Ultrafiltration is closely related to reverse osmosis with the distinction generally made on the basis of size of particle separated. Reverse osmosis removes all molecular sizes including inorganic salts. Ultrafiltration generally will not separate molecules smaller than ≈500-1000 MW, thus inorganic salts are not separated from solution. Ultrafiltration uses pressures of about 50 psi in contrast to reverse osmosis where pressures in excess of 500 psi are generally used.

Membrane ultrafiltration is a pressure activated process using semi-permeable membranes which act as molecular screens to separate molecular and colloidal materials dissolved or suspended in a liquid phase.

Thus far, the principal commercial applications of the process have been in (1) industrial operations where valuable products can be recovered by separation from a bulk solution, (2) analytical application which provides a new method to separate molecules according to size and molecular weight, and (3) in a package waste treatment plant which separates mixed-liquor solids from a biological reactor. Solids are returned to the aerator and clarified product water is discharged.

A plant employing ultrafiltration has been installed on Pikes Peak to provide waste treatment and water reuse. The essentials of the process are shown in Fig. 13. The 15,000 gpd plant treats wastewater by high-solids activated sludge. The solids are separated in an ultrafiltration unit. Solids are returned to the aerator. Product water is excellent. Some typical removals are shown below.

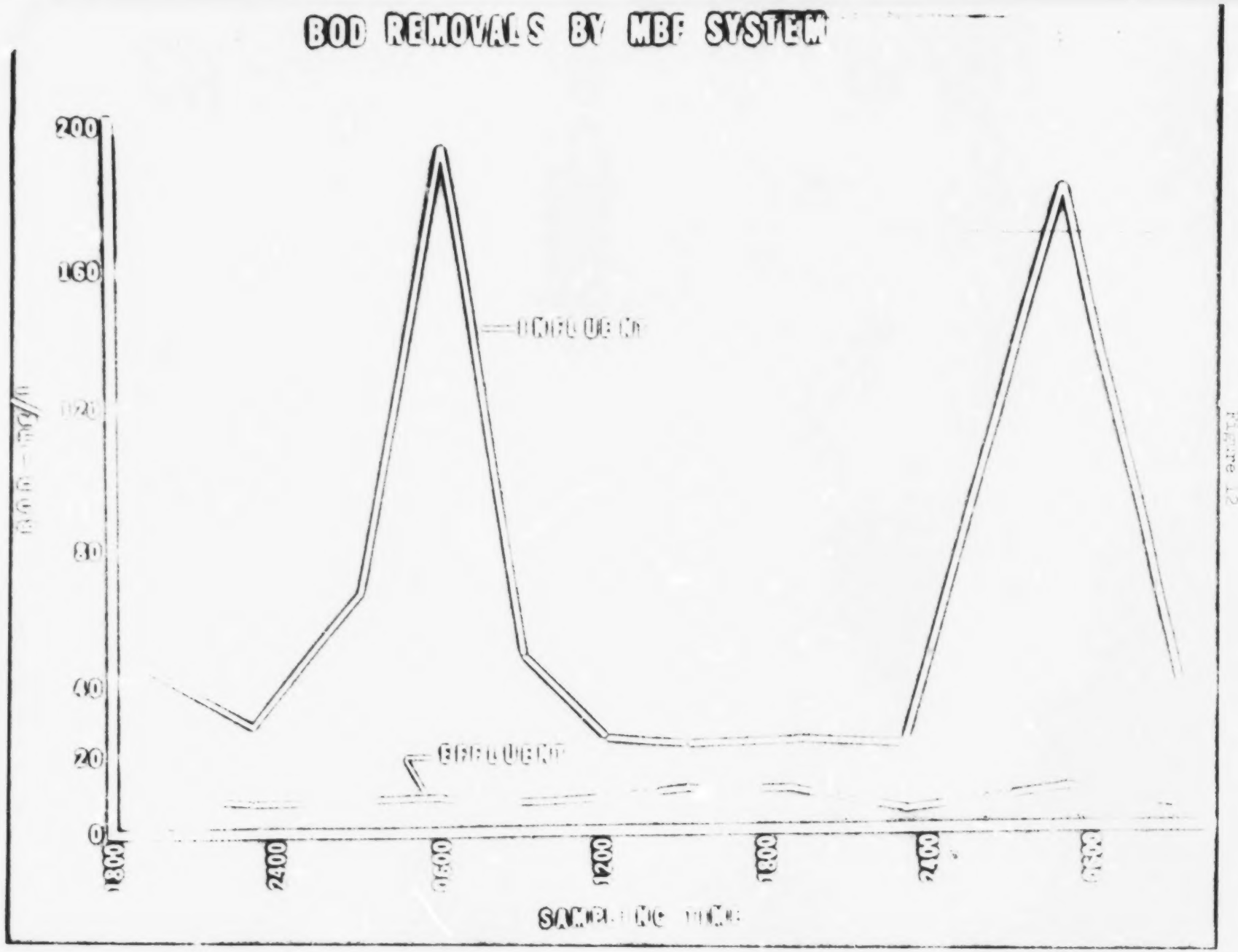

BOD REMOVALS BY MBF SYSTEM
200
160
120
80
40
20
0
1800
2400
0600
1200
0001
0072
0000

Figure 12

ENZYME REACTOR

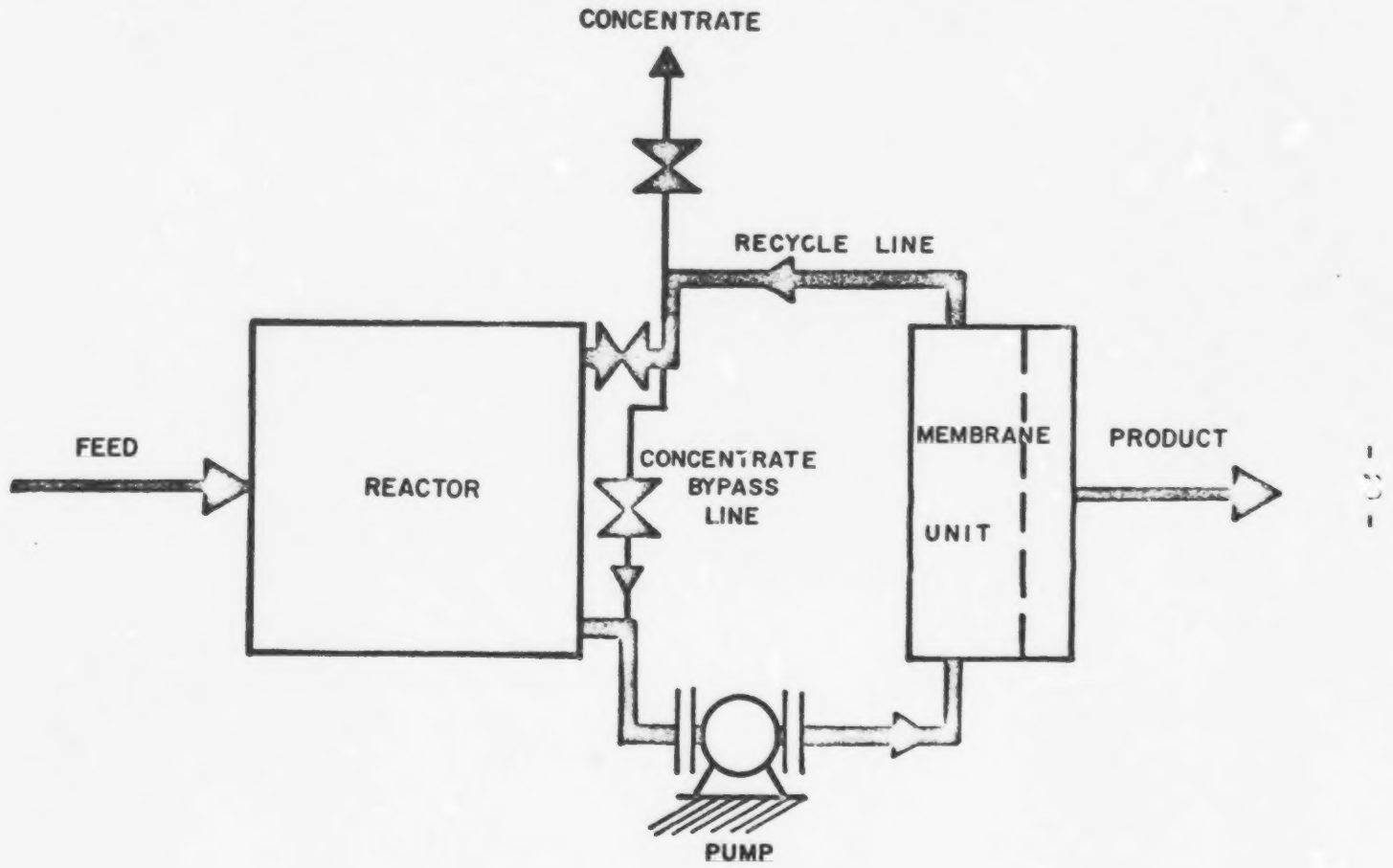


FIG. 13 REACTION SYSTEM

Removals by Activated Sludge and Ultrafiltration

	Influent	Effluent	% Removal
BOD	382	<1	>99
COD	678	20	97
Turbidity	-	<0.1	-
SS	323	0	100
PO_4-P	12.2	7.7	37

These results emphasize that only molecules or particles greater than 500-1000 MW are separated. Inorganic ions such as phosphate are not retained. The 37% reduction of phosphate was due to biological uptake in cells which were removed.

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NEW DEVELOPMENTS IN SLUDGE HANDLING AND DISPOSAL

Robert B. Dean

I. INTRODUCTION

A. Pollutants removed from wastewaters must be treated in such a way that they will not pollute the environment

1. Only three places to put polluting substances. Air, land, ocean. Not in surface waters.

B. Ocean Disposal

1. Ocean can oxidize wastes if kept aerobic.

a. Grease should be removed by aerobic treatment.

2. Food chains may concentrate poisons killing larger species.

C. Land Disposal. Fill.

1. Not suitable for soluble substances such as salts.
2. Needs dewatering to produce solid that will bear a load. Useful for insoluble inorganic wastes.

D. Land Disposal. Surface.

1. Not suitable for solubles except nutrients in quantities utilized by plants.
2. Low cost dewatering; can handle liquid sludges.
3. Low cost oxidation of organic matter.

E. Disposal into Wells

1. Unsuitable if there is a chance that pollutants will reach useful ground water or cause earthquakes.

II. ORGANIC SUBSTANCES

A. Principal problem in disposal of organic sludges is water. Twenty to fifty times as much water as all other substances in waste.

1. Oxidation of 1 lb. of organic sludges is sufficient to evaporate about 2 lb. of water.

2. Economics of incineration are therefore closely tied to dewatering by sedimentation, filtration and drying.

B. Treatment of wet sludge to aid further processing

1. Anaerobic digestion. Reduces solids about 50% by hydrolysis and fermentation to methane gas which is burned to CO_2 and water. Produces foul supernatant liquor which returns organics and nutrients to the plant for recycle.
2. Sludge cooking at 370°F. Improves filterability of solids. Returns 10-20% of the BOD and 60-80% of the nitrogen.
3. Wet oxidation - Zimpro at 350°F removes 15% of COD by oxidation; dissolves 25% of solids and 90% of nitrogen. Higher temperatures destroy more solids. Improves filtration, produces a foul supernatant liquor.
4. Aerobic Stabilization.

   a. Stabilizes solids, oxidizes oil and grease, retains nutrients.
   b. Can reduce nitrogen levels by nitrification-denitrification.
   c. Does not aid dewatering.

C. Oxidation and dewatering on land surfaces. An "old-fashioned" process.

1. Low cost dewatering on land.
2. Ammonia, phosphate and pesticides fixed on soil minerals.
3. Soil organisms destroy pollutants.

   a. Organics go to CO_2.
   b. Ammonia converted to nitrates.

4. Soil improved for agriculture and forestry.

III. CHEMICAL SLUDGES

A. Lime

1. $CaCO_3$, $Ca_5OH(PO_4)_3$, $Mg(OH)_2$, $CaSO_4$, $Ca(OH)_2$ Dewater in lagoon or vacuum filter.
2. Land fill
   a. Alkaline due to excess $Ca(OH)_2$
   b. Non-leachable $PO_4^{\equiv}$
   c. $SO_4^{=}$ leaches out, 1300 mg/l
3. Recovery of CaO from $CaCO_3$ by burning.

B. Iron

1. $Fe(OH)_3$, Fe^{+2}, $2\ Fe^{+3}\ (OH)_8$, $FePO_4$, $CaCO_3$, $CaSO_4$ -- Dewatering more difficult.
2. Land fill.
3. Iron recovery rarely worthwhile.

C. Aluminum

1. $Al(OH)_3$, $Al\ PO_4$, $CaCO_3$.
2. Very difficult to dewater; Freezing.
3. Aluminate recovery may be practical.

IV. SPECIAL DEWATERING TECHNIQUES

A. Freezing -- Destroys colloidal structure; gives good filtration

1. Needs slow freezing over several hours after thawing.
2. Complete freezing and minimal mechanical agitation.
3. Most promise for Activated Sludge and Aluminum Hydroxide.
4. Does not increase strength of filtrate.
5. Does not kill organisms.

B. Radiation

1. Selectively attacks high polymers and may reduce water holding capacity.
2. Does not disinfect. May kill 90% of microorganisms.
3. Releases organics to filtrate.
4. Requires costly shielding.

C. Pressure Cooking -- 250° C (480° F)

1. Gives good filterability.
2. Strong "soup" needs further treatment.
3. Disinfects.

D. Cooking with SO$_2$ Catalyst

1. Good filterability at lower temperatures and pressures.
2. Strong "soup" may have food value for livestock.
3. Disinfects.

E. Cooking with Oxygen -- Zimpro

1. High temperature to destroy organic solids (450° C)
2. Low temperature to get good filterability (250° C)
3. Produces "soup" that is difficult to treat.
4. Disinfects.
5. Power recovery not economical.

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REMOVAL OF ORGANICS FROM WASTEWATER BY ACTIVATED CARBON

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SUMMARY

A 0.3 mgd, four-stage, fixed bed granular activated carbon pilot plant has continuously treated unfiltered activated sludge plant effluent from the Pomona Water Renovation Plant from June, 1965, through July, 1969. Extensive operating data collected during four adsorption cycles, which required four years to complete, indicated that a high quality product water, characterized by an average dissolved COD of 8 mg/l was produced on a routine basis. This water contained, on the average, a concentration of carbon chloroform extract (CCE) equivalent to 0.026 mg/l, which is significantly less than that required by the Public Health Service standards for potable water.

Successful backwashing of the first stage carbon, which serves as a filter and an adsorbor, made pretreatment of the secondary effluent unnecessary. The filtering action of the carbon readily removed the 9 mg/l of suspended solids present in the activated sludge plant effluent. This filtration causes, on the average, a 10 psi increase in the pressure loss over a 24 hour service period. Partially due to filtration, 46 percent of the total COD that is removed by the four stage column is removed in the first stage. About equal removals of dissolved organic matter, represented by dissolved COD, occur in each stage.

Removal efficiency of dissolved COD was greater than anticipated, amounting to a carbon capacity of 59 pounds per 100 pounds of carbon during the first cycle. A 17 percent decrease in the carbon capacity occurred

between the first and fourth cycles which resulted in a fourth cycle capacity of 49 pounds of dissolved COD per 100 pounds of carbon. The data indicates that a steady state condition is being reached at a capacity of 46 to 50 pounds per 100 pounds of carbon.

Routine column performance data show that 89 percent of the secondary effluent dissolved COD was removed during the first cycle, which decreased to 71 percent during the second cycle after which the removal approached a steady state condition. Following an increase in the carbon dosage from 250 pounds of carbon per million gallons during the first cycle, the dosage remained constant during both the third and fourth cycles at 350 pounds per million gallons.

Regeneration of the carbon was shown to be a feasible process and has been routinely accomplished. Iodine number determinations confirm that a decrease in adsorptive capacity takes place and that the micropores (less than 20 A^{o}) in the carbon cannot be completely cleared of adsorbed organics by thermal regeneration. However, the iodine molecule is not characteristic of typical organic matter in secondary effluent and consequently, as the data show, is not a satisfactory parameter for determining the effect of repeated adsorption cycles on the capacity of carbon for dissolved COD.

Data collected from a separate, parallel, single stage column designed to provide supplementary information to that collected from the four-stage column indicated a 13 percent decrease in the mean particle diameter of the carbon after seven adsorption cycles at which time a steady state condition existed. No significant differences in column operation or

in pressure loss in the carbon resulted due to the decrease in particle size from 0.90 to 0.79 mm.

An estimate for a 10 mgd, four-stage carbon pilot plant designed to produce a product water with a dissolved COD of 8 mg/l was made. The estimated cost of $86 per million gallons is based on a fourth cycle carbon dosage of 350 pounds per million gallons of water and a conservative carbon loss of 10 percent per cycle.

INTRODUCTION

The Federal Water Pollution Control Administration and the Los Angeles County Sanitation Districts, through a contractual agreement, operate an advanced waste treatment research facility on the grounds of the Pomona Water Renovation Plant in Pomona, California. The over-all objective at Pomona is the development of the minimum cost technology capable of renovating wastewater. In order to attain this objective, the most promising available treatment processes are being field tested and evaluated on a pilot plant scale.

The removal of residual dissolved and suspended organic matter from municipal waste water treatment plant effluents is a necessary step if this water is to be reused. It has been demonstrated at Pomona that the use of a granular activated carbon column is a feasible process for accomplishing this removal.

Since June of 1965, an 0.3 mgd, four-stage, fixed bed, granular activated carbon pilot plant, complete with thermal regeneration facilities, has been continuously treating unfiltered activated sludge plant effluent from the Pomona Water Renovation Plant. The objective is to obtain reliable cost and operating data on the use of granular carbon and to establish its effectiveness, after repeated adsorption cycles, in removing dissolved and suspended organic matter that vary in type and concentration due to the conditions of daily and seasonal flow fluctuations that occur in municipal waste streams.

PILOT PLANT DESCRIPTION

The pilot plant components include five carbon vessels or contactors, influent and product water storage tanks, a carbon dewatering bin, a quench tank, and a regeneration furnace, all mounted on a concrete slab with related piping, pumps and controls. The slab is 44 feet square and sloped to drain into the quench tank pit in order to facilitate the recovery of any accidentally spilled carbon. A photograph of the carbon pilot plant is presented in Figure 1. (omitted from this seminar paper)

A schematic flow diagram for the pilot plant is shown in Figure 2. Initially, chlorinated secondary effluent was used since it was more conveniently available from the water renovation plant. However, during the course of the study the plant was enlarged by the construction of a new parallel treatment plant and both chlorinated and unchlorinated secondary effluent from either the old or new treatment plants has been used without any significant differences in operation of the carbon columns.

The secondary effluent is directed to the storage tank from where it is pumped into the carbon columns. The pilot plant is composed of two columns, a single stage and a four stage; each of the stages or contactors is identical in design and contains 6,700 pounds of 16 x 40 mesh granular carbon having characteristics as shown in Table I. The contactors are 6 feet in diameter, have an overall height of 16 feet, and are designed for a working pressure of 50 psi. The detailed design is shown in Figure 3. The carbon bed is supported on a Neva-Clog screen system consisting of the

Neva-Clog screen, an expanded metal backing, and a series of concentric annular rings which are an integral part of the contactors for structural support. The screen is comprised of two layers of stainless steel perforated with 1/32" diameter holes and spot-welded together with holes off-set from each other.

The interiors of the contactors are coated with a coal-tar epoxy compound designed to withstand the corrosive action of the carbon-water environment. One-inch couplings are welded in the contactor walls for sampler connections and pressure taps.

The liquid enters at the top from an annular ring through twenty 1" holes around the circumference of each contactor. This ring is also used to remove liquid during backwashing. A distance of 3-1/2 feet between the entrance ring and the top of the carbon bed provides room for a 35% to 40% bed expansion during backwashing.

The single stage column was designed to provide supplementary data to that obtained from the main four stage column pilot plant and is operated at 200 gpm, or a hydraulic loading of 7 gpm/ft^2, which results in an empty bed detention time of 10 minutes. The water leaves the column and is discharged to the nearby San Jose Creek.

The four stage column also operates at 200 gpm and has an empty bed detention time of 40 minutes with the secondary effluent passing, in series, through each of the stages and thence to the product water storage tank with overflow entering the creek. As the carbon becomes exhausted, the level of organic matter in the effluent increases and, when it reaches a COD of 12 mg/l, the first stage is removed from service and the carbon is

transferred as a slurry into the [illegible] tank. The carbon, with a 40 percent moisture content, is conveyed to the multiple hearth furnace where it is regenerated at [illegible].

The furnace is a 30 inch I.D., vertical, refractory-lined, 6 hearth unit, complete with air and [illegible] equipment. It is natural gas fired and steam is added to enhance the regeneration. Carbon leaves the furnace at 90 lbs/hr and is [illegible] under water, into the quench tank from where it is educted into the contactor. Regeneration of one stage requires about 75 hours to complete.

Carbon losses during regeneration, [illegible] average 5-1/2 percent, are made up with virgin carbon and the [illegible] is returned to service as the fourth stage of the column. Simultaneously the other stages are moved forward one position, thus maintaining a counter-current flow of carbon and water.

Repeated exhaustions and regenerations of the carbon are defined as adsorption cycles. In the four stage [illegible] adsorption cycle for the column is completed when all of the stages have been exhausted and regenerated once, that is, each [illegible] in service in each stage position. Experience has indicated that regeneration of the first stage is required every 75 days--equivalent to treating about [illegible] of secondary effluent. On this basis, exhaustion of each of the four stages requires almost one year. At this rate four cycles were not completed during the period of this study.

In the single stage column [illegible] cycles are completed in a year's time. This has allowed evaluation of the effect of 9 cycles on

carbon efficiency and 11 cycles on carbon attrition and regeneration characteristics. During the last two cycles, special studies were conducted with secondary effluent which did not include the monitoring of dissolved COD removal, but did require carbon regeneration. Though its operation is different from that of the main four-stage pilot plant, the single stage column has provided valuable additional supporting data.

PILOT PLANT PERFORMANCE

Operation

Backwashing of both the single stage column and the first stage of the four stage column is performed on a daily basis to free the carbon bed of trapped suspended solids removed by filtration from the secondary effluent. Backwashing is accomplished using secondary effluent at hydraulic loading rates up to 7 gpm/ft^2 and a volume of water equivalent to 3 to 4% of the daily column flow.

Pressure loss data in the carbon for each stage of the four stage column, before and after backwashing the first stage, is shown in Table 2. After backwashing the first stage, the column has a total pressure drop of about 21 psi which increases to 32 psi over an average 24 hour period. This increase, as presented in Table 2, is due to filtration of suspended solids by the first stage carbon resulting in a 10 psi pressure loss increase in this stage.

The second stage must also be backwashed to maintain the column operating pressures within the capacity of the pumping system. This is required, on the average, once or twice per month.

The use of granular activated carbon in a column system requires that it be regenerated, transported to and from the regeneration facilities, and routinely backwashed. As a result of those operations, attrition of the carbon particles occurs. Data showing the effects of carbon regeneration and handling on the mean particle diameter is presented in Figure 4. A 13

percent decrease in the particle diameter of the carbon in the single stage column occurred after 7 adsorption cycles, at which time no further significant changes in the size took place. This is primarily caused by the addition to the column of virgin make-up carbon, which has a mean particle diameter of 0.9 mm, after each regeneration to replace losses. Also, as the supply of the original 16 x 40 mesh carbon was depleted, Filtrasorb 400 (12 x 40 mesh carbon having a mean particle diameter of about 1 mm) was purchased and used as make-up after the seventh cycle.

In addition to the reduction in the carbon particle size previously described, which has not cuased any significant difference in the carbon column operation or pressure drop, a loss of carbon occurs during the regeneration process. This loss has ranged from 11 percent during the first cycle to 7 percent during the fourth and has averaged 8-1/2 percent during the four year study.

Adsorptive Capacity

The COD test has been used as the primary parameter for determining the efficiency of the carbon columns in removing both suspended and dissolved organic matter. Since carbon serves as a filtration media as well as an adsorbant, the COD of the secondary effluent and carbon effluent was determined on both unfiltered and filtered (0.45 micron, membrane filter) samples and values for total COD and dissolved COD (filtered) are reported.

The data in Figure 5 show the distribution of COD removal in the four stage column. The dissolved COD data indicate about equivalent removal in each of the stages. However, the total COD data show 46 percent is removed in the first stage. This difference is due to filtration of suspended COD

which accounts, on the average, for 20 to 30 percent of the total COD. Therefore, when monitoring the effect of repeated adsorption cycles on the carbon efficiency it is necessary that a measurement of the dissolved organic matter be made. A carbon column even after its adsorptive capacity is exhausted will continue to remove suspended COD.

The effect of adsorption cycle on the dissolved COD removal performance of the four stage column is shown by the data in Figure 6. Each of the data points represents a regeneration of the first stage column; four such regenerations are equivalent to an adsorption cycle. The data clearly indicate that a reduction in COD removal occurs from an initial level of 89 percent to 71 percent after 1-1/2 adsorption cycles, and that the removal remains constant through the remaining cycles.

Comparable data for the single stage column are presented in Figure 7. The dissolved COD removal decreased from an initial value of 53 percent to 40 percent during the sixth cycle after which time the removal stabilized at the 40 percent level. The addition of 8-1/2% virgin make-up to replace losses during each cycle has had the effect of upgrading the adsorption capacity of the carbon. If this loss rate is decreased, the loss in capacity with adsorption cycle will be greater.

The effect of repeated adsorption cycles on the adsorptive capacity of carbon can be expressed in terms of its capacity for a given adsorbate under specific laboratory conditions. One of the adsorbates used to characterize the adsorptive properties of activated carbon is iodine. The standard iodine number test is used to measure the extent of recovery of the adsorptive capacity or the extent to which the micropores (less than 20 A^{o}) have been

cleared of adsorbed organics by regeneration. Data in Figure 8 show the iodine number for both spent (exhausted) and regenerated carbons from the single and four stage columns after repeated cycles.

The initial iodine number of the virgin carbon was 1100 and after eleven adsorption cycles in the single stage column the iodine number was reduced to 608 which is a 45 percent decrease with every indication that further reductions will take place. An even greater decrease with adsorption cycle is noted in the iodine number of the regenerated carbon in the four stage column. This continuous decrease in the adsorptive capacity of the carbon for iodine indicates the micropores have not been cleared of adsorbates during regeneration. However, previous data shown in Figures 6 and 7 relate COD removal to adsorption cycle and indicates a 20 percent reduction in efficiency in the four stage column during the first two cycles after which time the removal remains constant. Iodine is therefore not typical of the organic matter in secondary effluent and although the surface area in the micropores has continued to decrease, the capacity of the carbon for organic matter as measured by dissolved COD has not followed suit.

The iodine number of spent carbon as shown in Figure 8 also shows a steady decrease forth both single stage and four stage columns. The difference between the regenerated and spent iodine numbers, or iodine capacity used, is due to the presence of substances in the micropores that were removed from the water during operation of the column. The data indicate the amount of capacity used reaches a steady state condition after the third cycle.

Since carbon capacity can have a significant effect on the economical

feasibility of carbon adsorption as a process for removing organic matter from secondary effluent, it is necessary to ascertain the effect of repeated adsorption cycles on the virgin or original capacity. Data showing this effect are presented in Figure 9. The total COD capacity is somewhat variable and is primarily due to the variability of the suspended solids in the secondary effluent which accounts for 20 to 30 percent of the total COD. The dissolved COD in the secondary effluent, which is significantly less variable, shows a more consistent relationship between capacity and adsorption cycle. It has been previously mentioned that dissolved COD was chosen as the primary parameter to measure the change in adsorptive capacity. The capacity is expressed in pounds of COD adsorbed per 100 pounds of carbon and has decreased 17 percent after four adsorption cycles. The change in capacity between the third and fourth cycles has decreased less than that of the early cycles indicating a steady state condition is being reached at 45 to 50 lbs/100 lbs.

It has been established through operating experience at Pomona that biological oxidation of organic matter takes place in carbon. The single stage column during a special long term study was never completely exhausted and continued to remove 2 to 4 mg/l of dissolved COD at the end of a nine month period; for this reason biological action undoubtedly contributes to increasing the carbon capacity above that occurring from true adsorption phenomena.

The relationship between carbon dosage and adsorption cycle is shown in Figure 10 and indicates a steady state condition was reached after the third cycle at 350 lbs of carbon per million gallons.

Water Quality

The average quality characteristics of the four stage column influent and effluent from June, 1965, to July, 1969, are shown in Table 3. The total COD of the effluent has averaged 10 mg/l during this period with yearly averages ranging from 9.3 to 10.5 mg/l. In order to evaluate the performance of the column after repeated adsorption cycles it was necessary to maintain a constant effluent quality in order to eliminate this as a variable in determining carbon dosage and capacity.

The United States Public Health Service Drinking Water Standards for 1962 limit the concentration of carbon chloroform extract (CCE) in a water supply to 0.2 mg/l. The CCE present in the carbon effluent has averaged 0.026 mg/l, which is an order of magnitude less than that required for potable water.

The removal of turbidity and suspended solids is due to filtration by the carbon and takes place primarily in the first stage.

A limited quality of nitrate is removed in the column and, at an influent concentration of 0.1 mg/l (NO_3-N), amounts to a 19 percent reduction and is due to biological denitrification in the carbon.

Cost Estimate

A cost estimate for a 10 mgd, four-stage carbon plant based on the operating results obtained during the four year pilot plant study is detailed in Table 4. The capital equipment cost of 1.86×10^6 was amortized over 20 years at 5% interest and amounts to 48 percent of the total cost of $86 per million gallons for carbon treatment. This estimate is based on a plant that would provide product water having the quality characteristic

shown in Table 3. Cost and quality are inter-related variables and must be considered together. For example, if a product water with an average total COD of 15 mg/l was desired it could be provided at a cost less than $86 per million gallons.

The cost of make-up carbon was determined using a dosage of 350 pounds of carbon per million gallons of water as shown in Figure 10. A conservative loss of 10 percent was also applied to estimate a makeup carbon cost of $11 per million gallons.

TABLE 1

CARBON CHARACTERISTICS

IODINE NUMBER	1100
METHYLENE BLUE NUMBER	300
MOLASSES NUMBER	220
APPARENT DENSITY	0.47 g/ml
ASH CONTENT	5%
MEAN PARTICLE DIAMETER	0.90 mm
MESH SIZE	16 x 40

TABLE 2

EFFECT OF STAGE A BACKWASH ON PRESSURE LOSS

	PRESSURE LOSS (PSI)				
	STAGE A	STAGE B	STAGE C	STAGE D	TOTAL
BEFORE BACKWASH	16.8	5.2	4.0	5.6	31.6
AFTER BACKWASH	6.6	4.9	4.0	5.2	20.7

TABLE 3

CARBON ADSORPTION PILOT PLANT

AVERAGE WATER QUALITY CHARACTERISTICS

JUNE 1965 to JULY 1969

PARAMETER		INFLUENT	EFFLUENT
SUSPENDED SOLIDS	mg/l	9	0.6
COD	mg/l	43	10
DISSOLVED COD	mg/l	30	8
TOC	mg/l	12	3
NITRATE as N	mg/l	8.1	6.6
TURBIDITY (Jtu)		8.2	1.2
COLOR (Platinum-Cobalt)		28	3
ODOR (Ton)		12	1
CCE	mg/l	-	0.026
BOD	mg/l	3	1

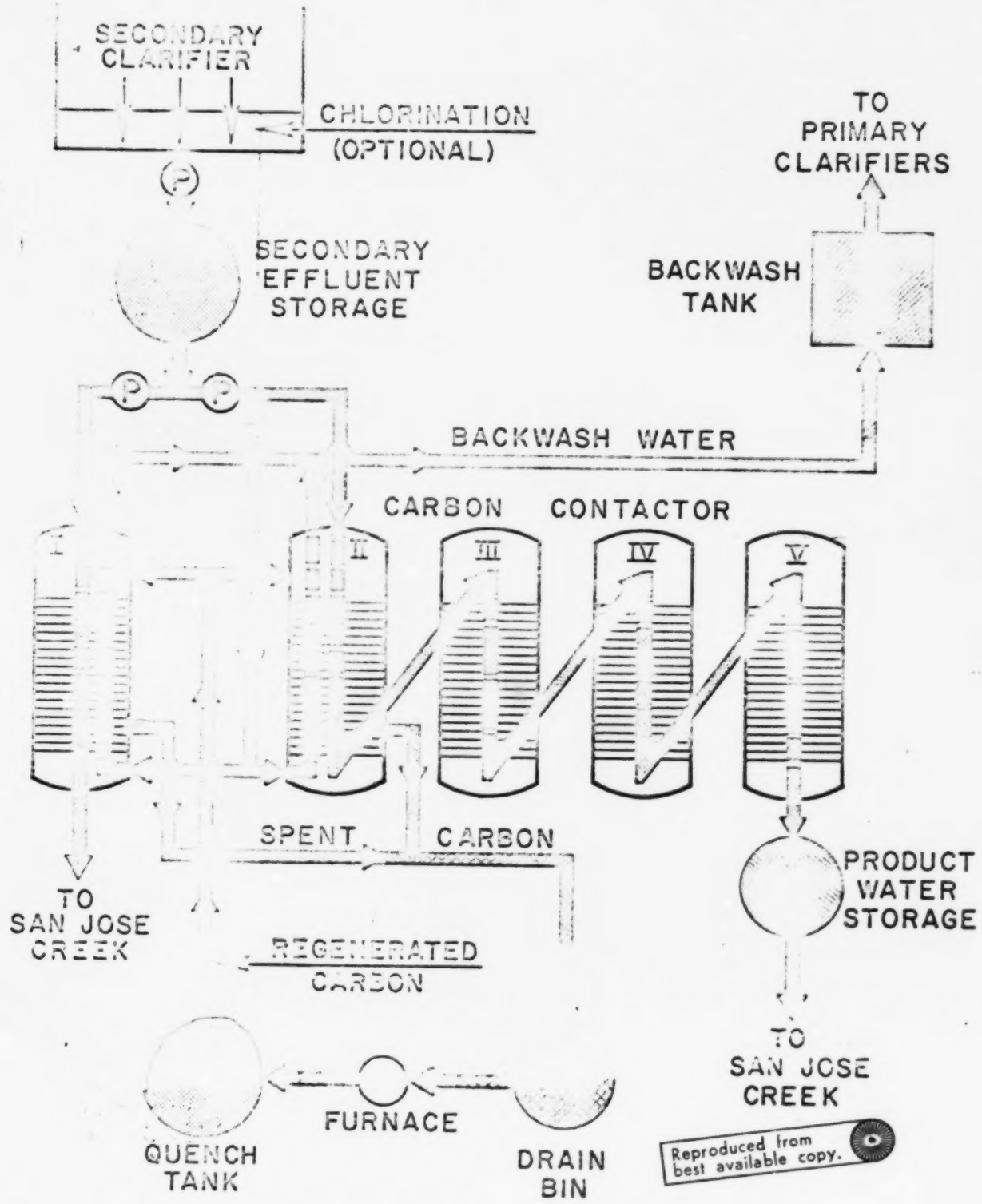
TABLE 4

CARBON ADSORPTION PROCESS

COST ESTIMATE - 10 MGD PLANT

AMORTIZATION OF CAPITAL		$/10^6 GALLONS
$1,860,000; 20 years @ 5%		41.00
OPERATION AND MAINTENANCE		
POWER	$ 8.50	
LABOR	18.00	
MAINTENANCE MATERIALS	5.00	
CARBON REGENERATION		
POWER, GAS AND WATER	2.50	
MAKEUP CARBON (10% Loss)	11.00	
		45.00
TOTAL		$86.00

SCHEMATIC FLOW DIAGRAM



Reproduced from best available copy.

TYPICAL CONTACTOR

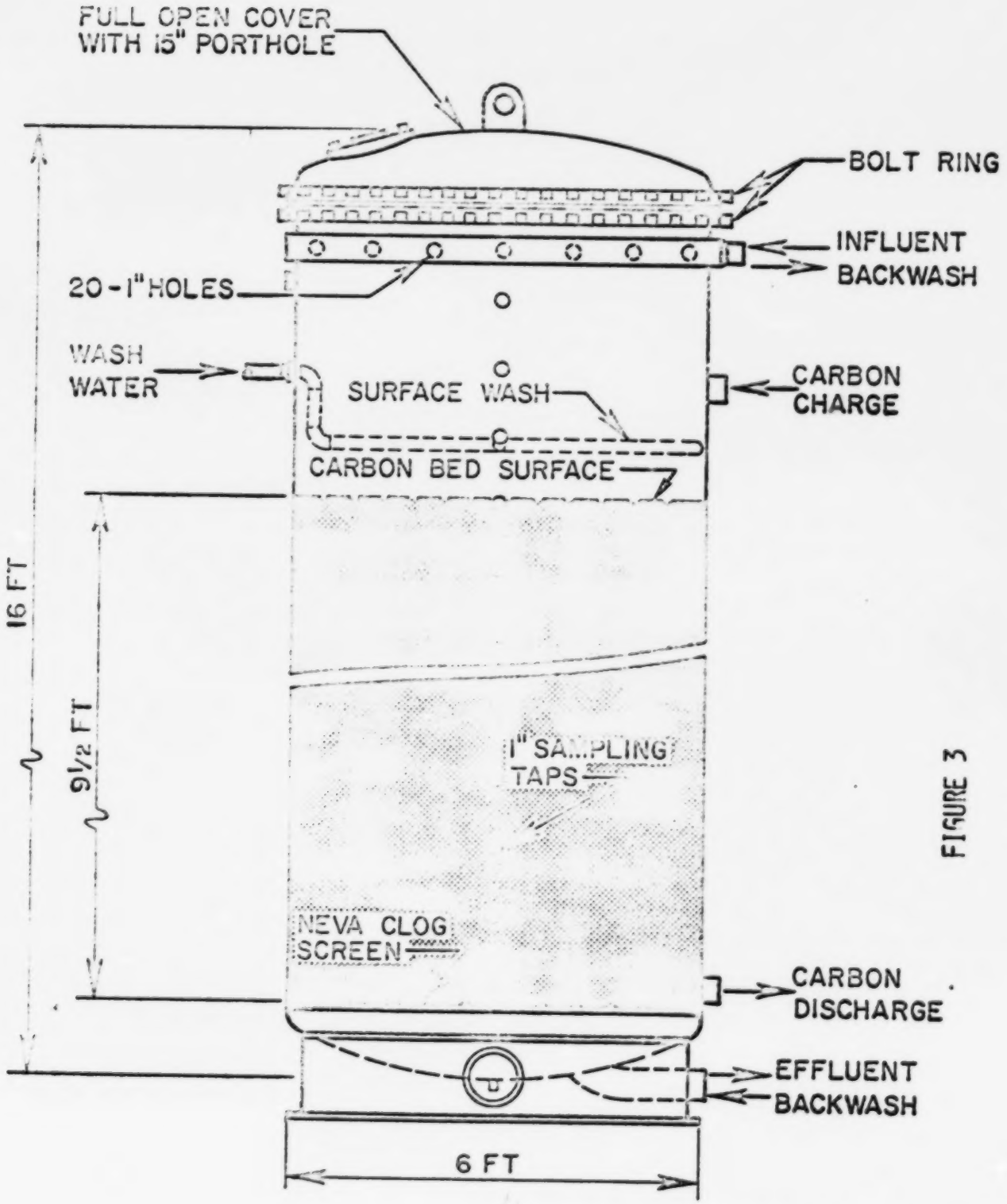


FIGURE 3

FIGURE 4

EFFECT OF ADSORPTION CYCLE ON CARBON PARTICLE SIZE

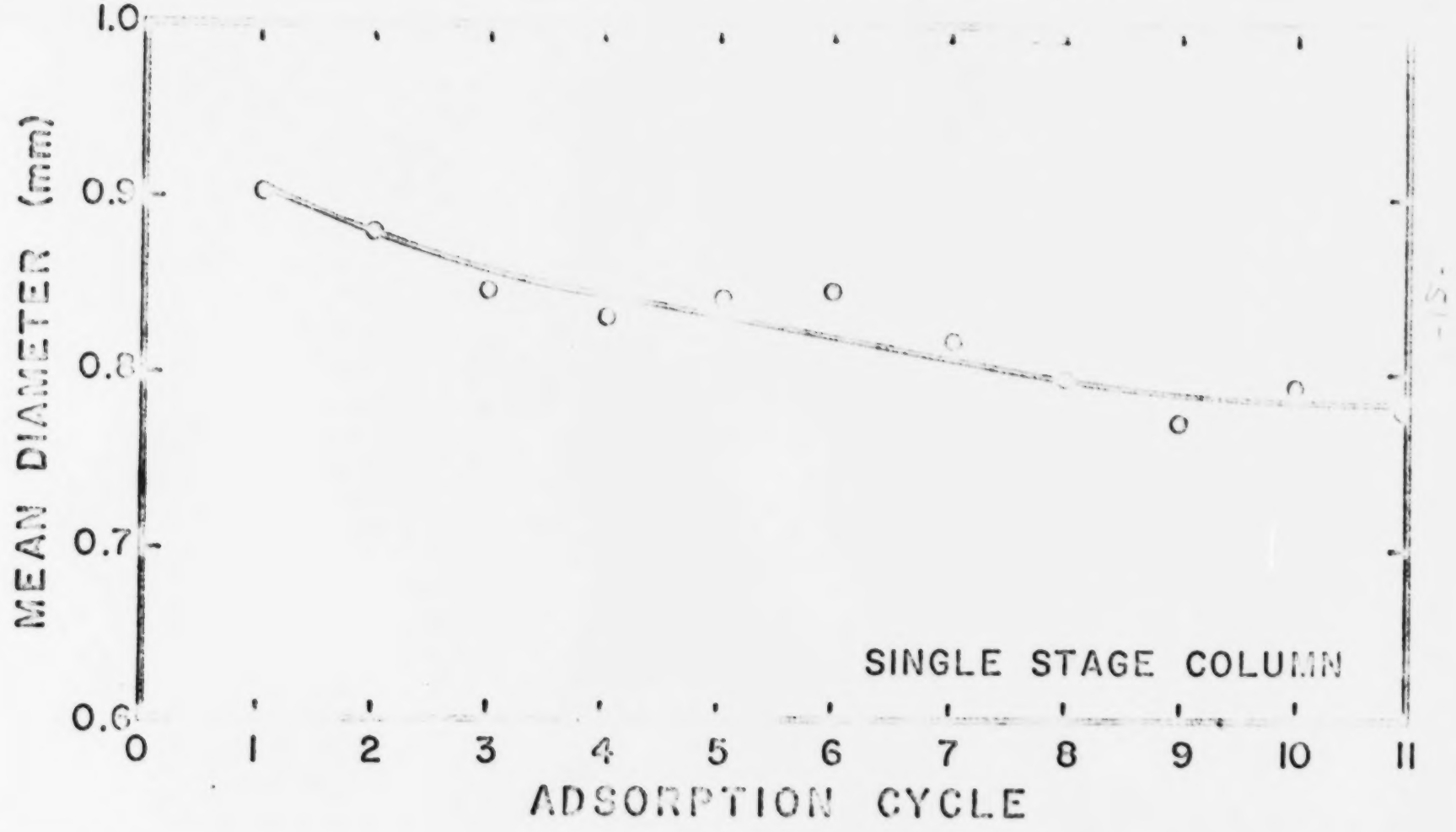


FIGURE 5

DISTRIBUTION OF COD REMOVAL

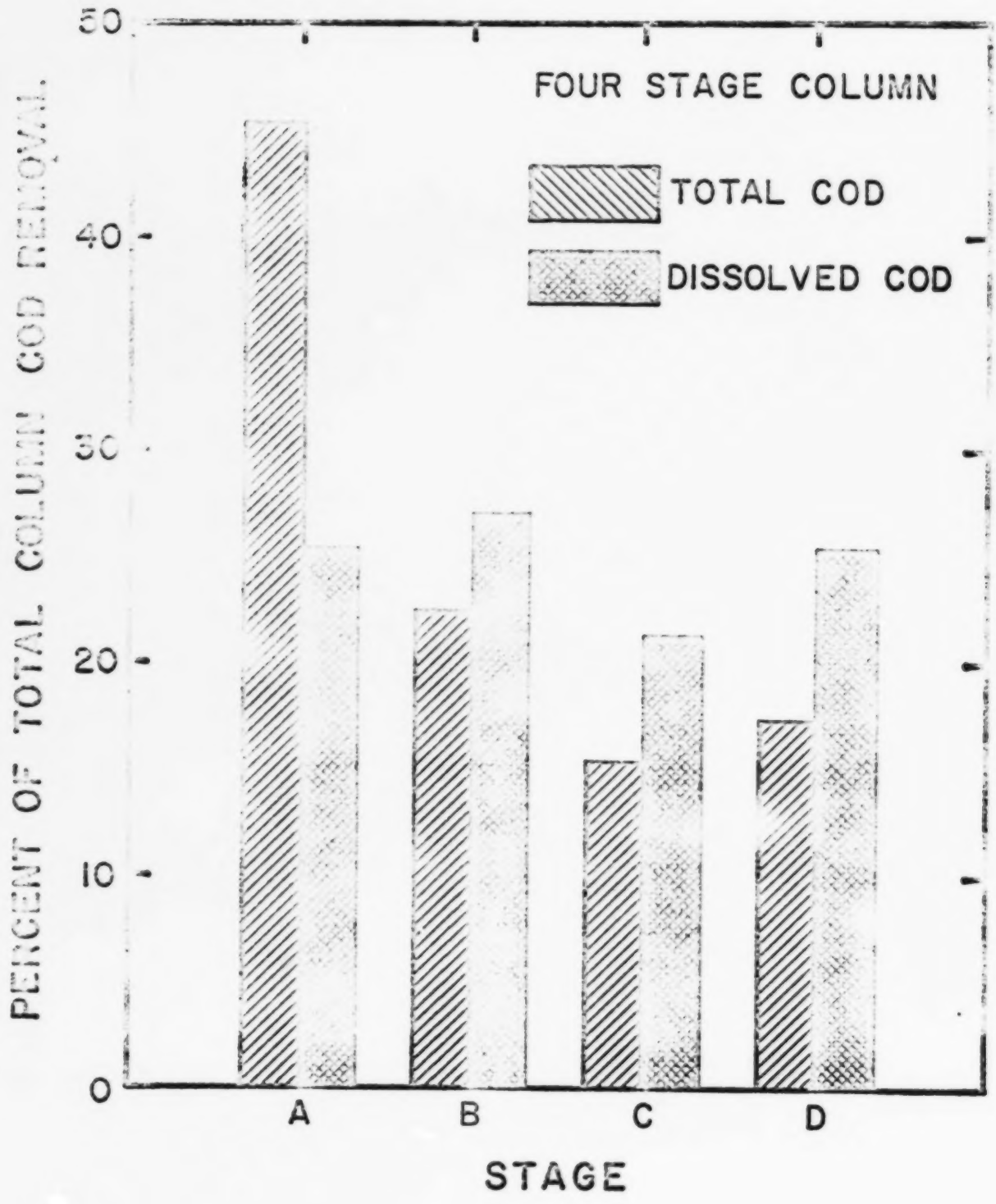


FIGURE 6

EFFECT OF ADSORPTION CYCLE ON COD REMOVAL

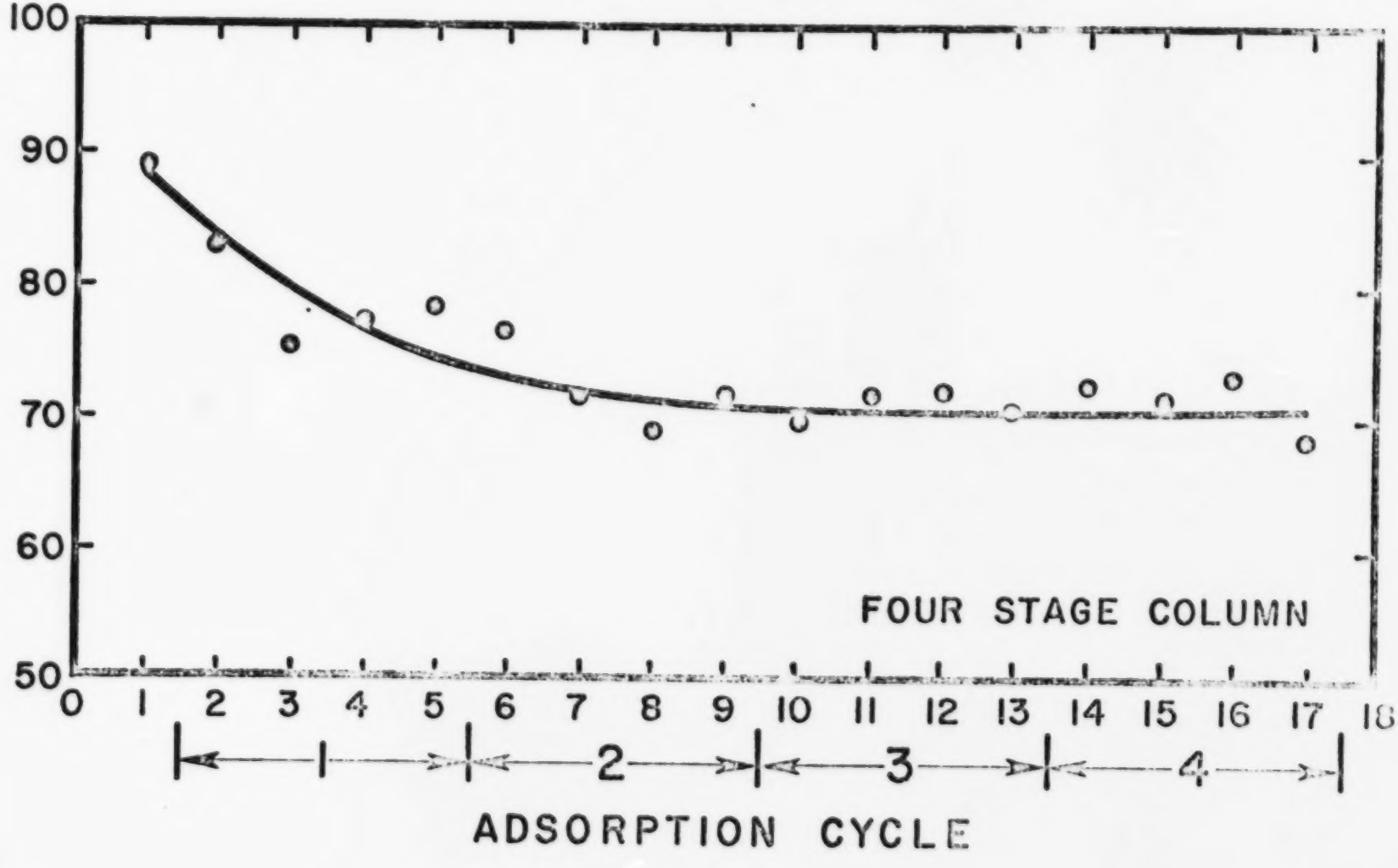


FIGURE 7 -

EFFECT OF ADSORPTION CYCLE ON COD REMOVAL.

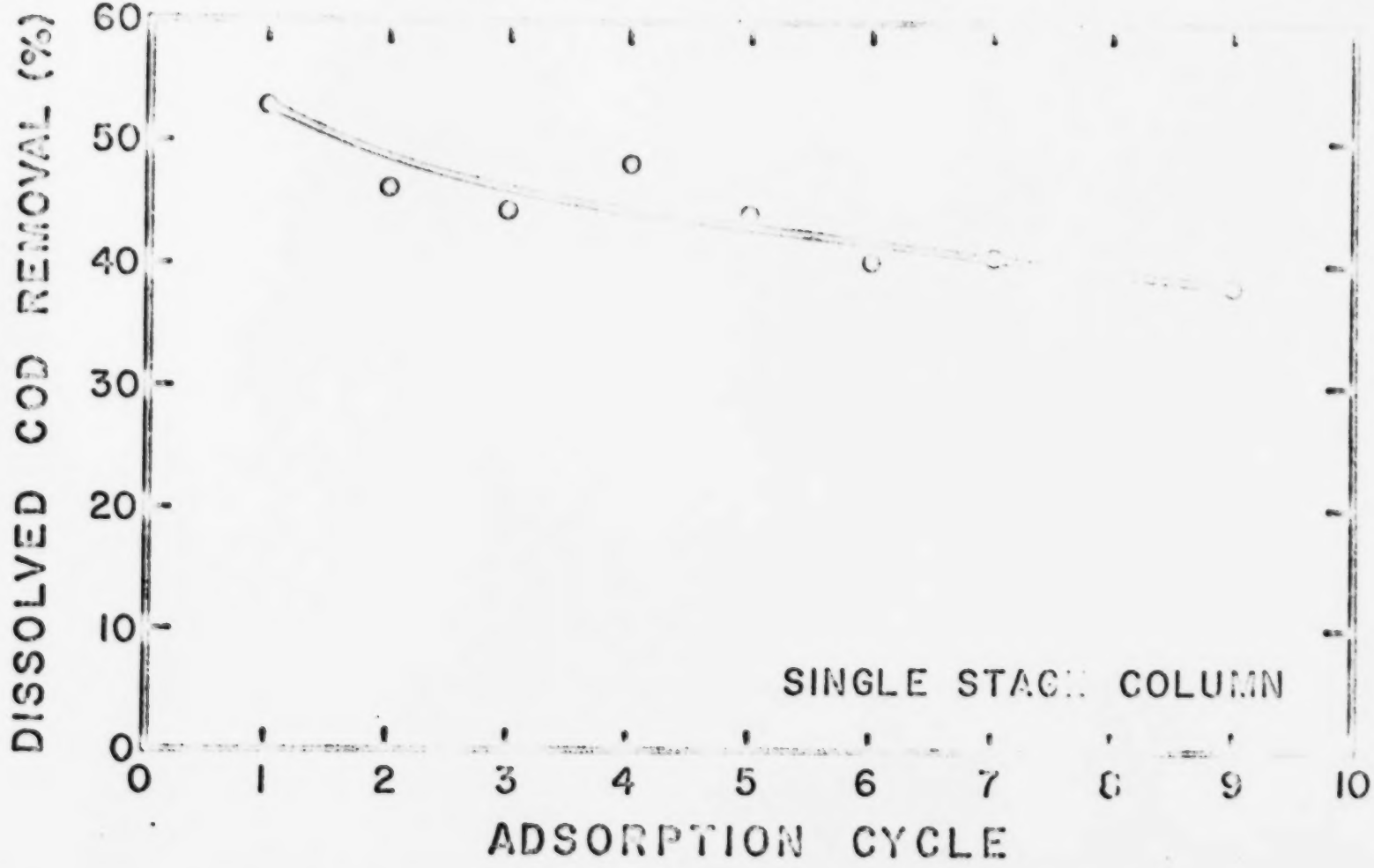


FIGURE 8

EFFECT OF ADSORPTION CYCLE ON IODINE NUMBER

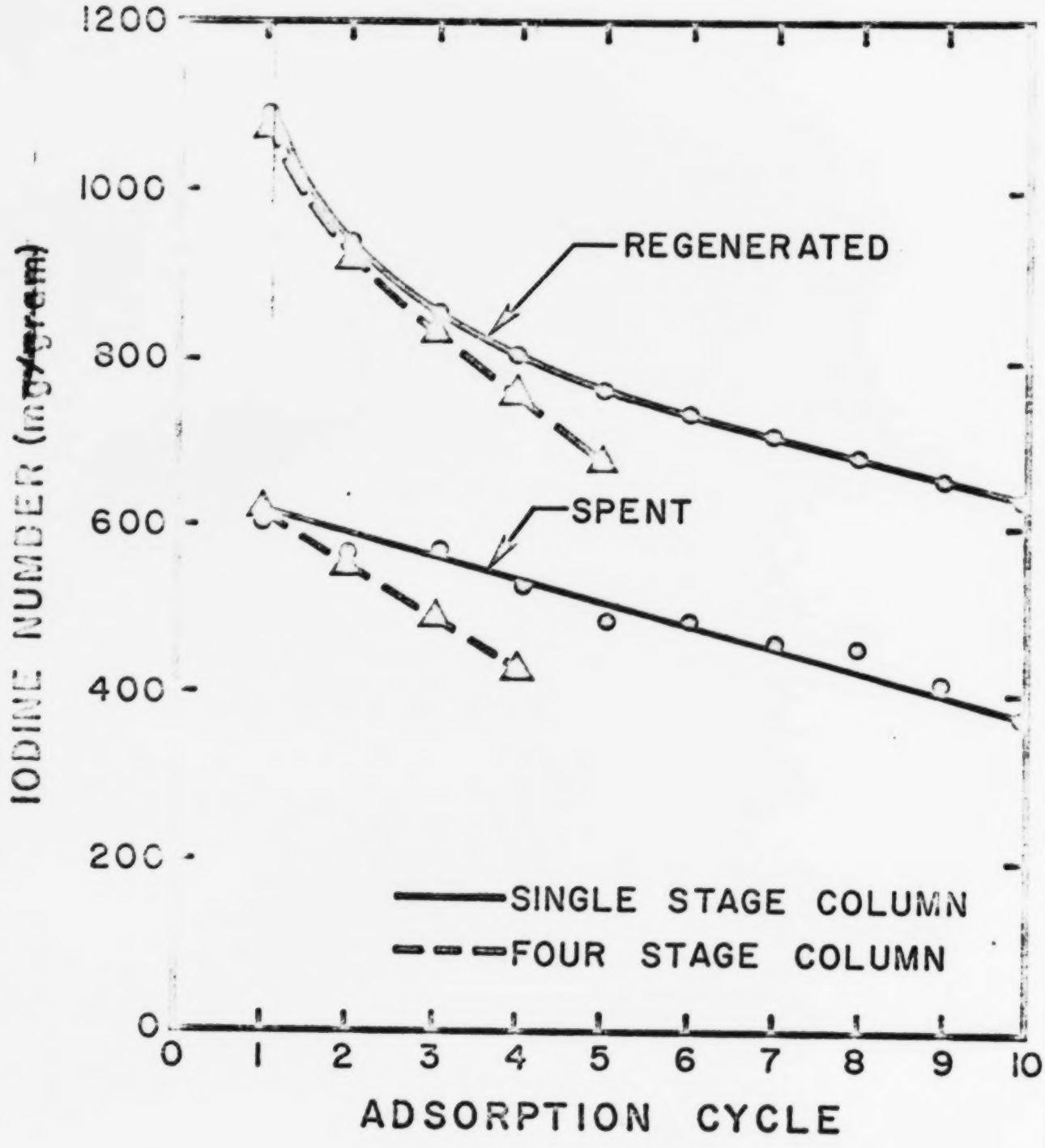


FIGURE 9

EFFECT OF ADSORPTION CYCLE ON CARBON CAPACITY

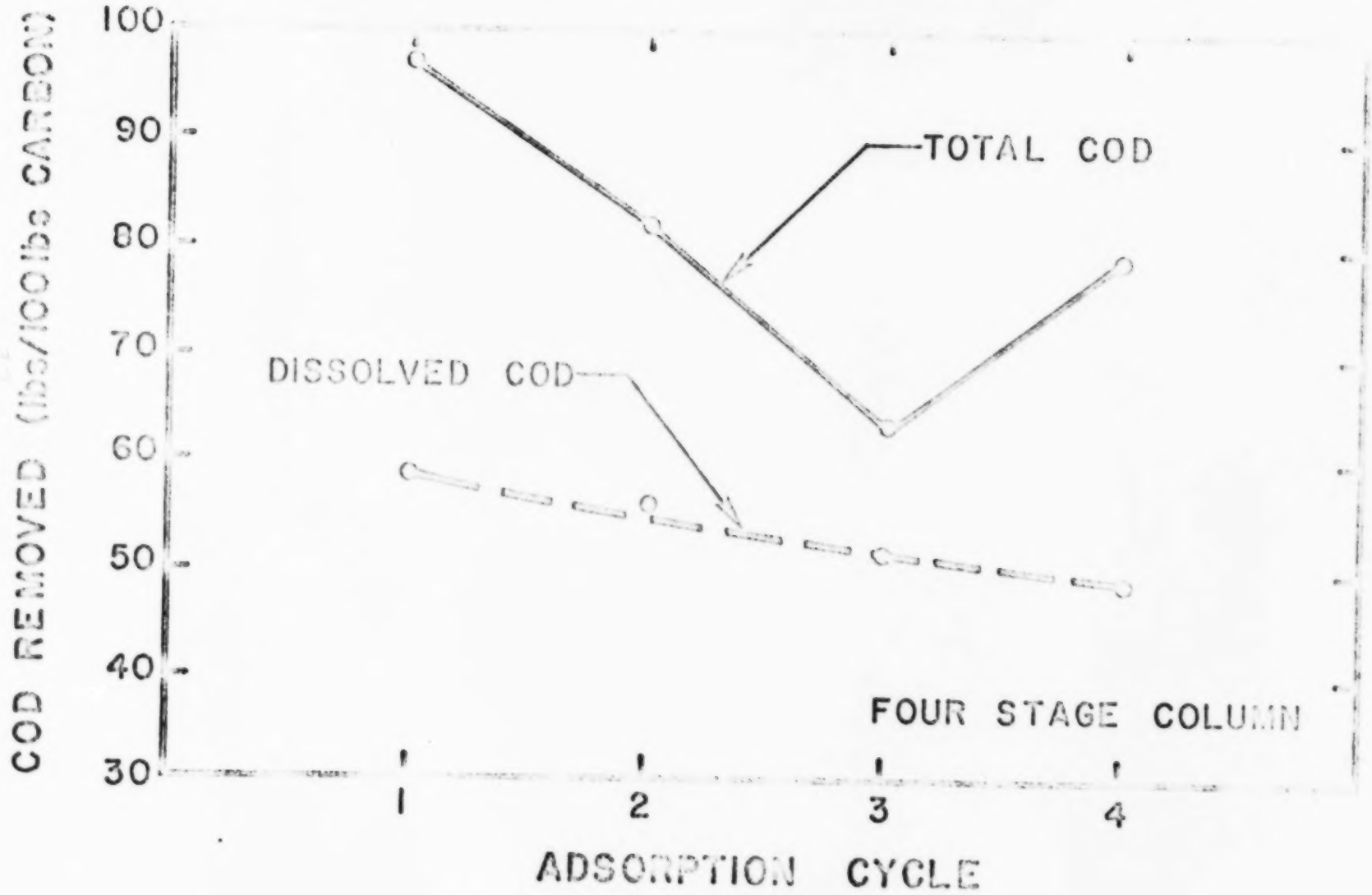
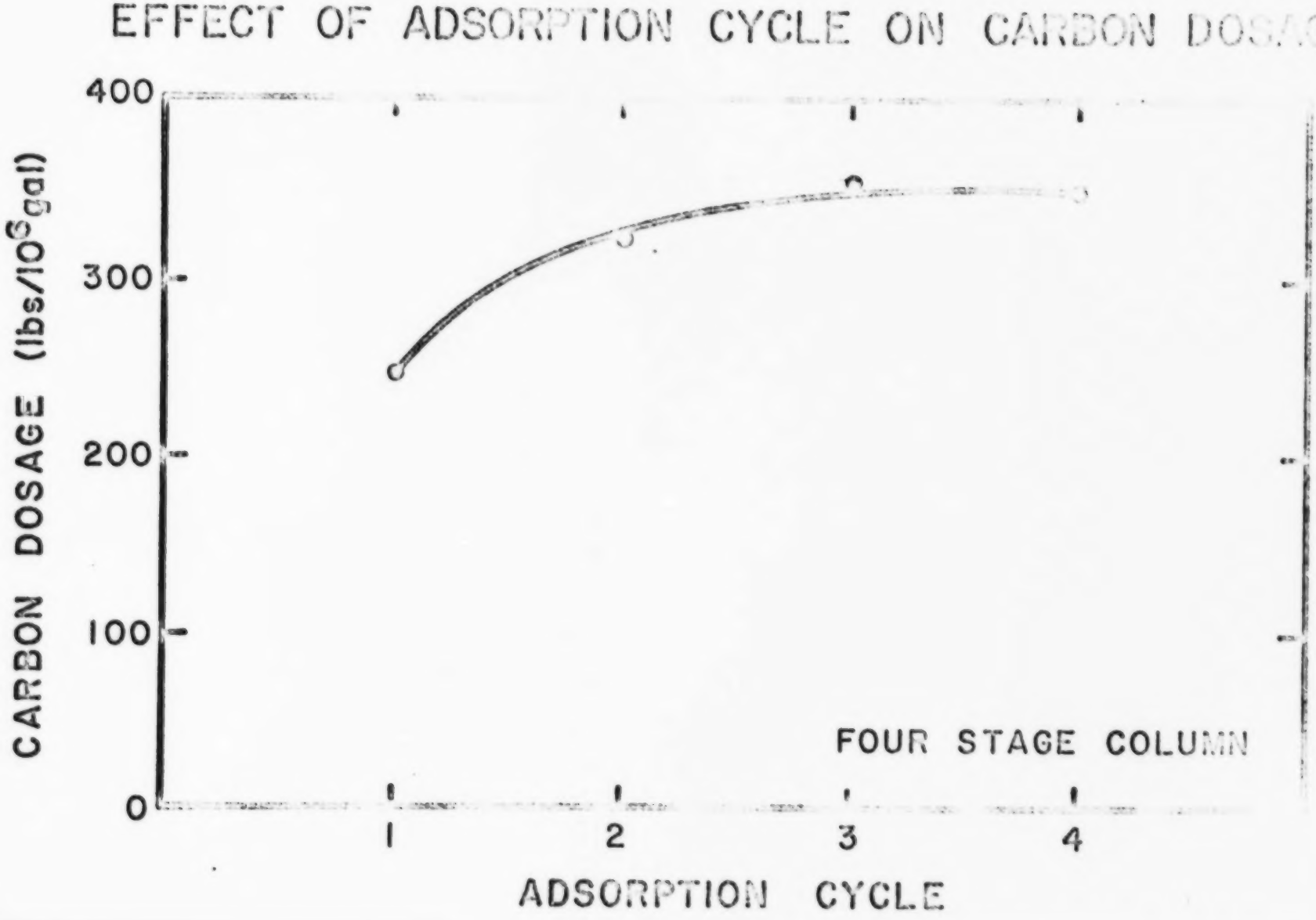


FIGURE 40


EFFECT OF ADSORPTION CYCLE ON CARBON DOSAGE
CARBON DOSAGE (lbs/10^6 gal)
400
300
200
100
0
1
2
3
4
ADSORPTION CYCLE
FOUR STAGE COLUMN

WATER POLLUTION CONTROL RESEARCH SERIES 17060DNW05/70

INVESTIGATION OF THE USE OF HIGH PURITY OXYGEN AERATION IN THE CONVENTIONAL ACTIVATED SLUDGE PROCESS

Excerpt of Report

ABSTRACT

A full scale system designed to demonstrate the practical and economical use of high purity oxygen gas in aeration of the conventional activated sludge waste treatment process has been tested under a variety of operating conditions in a direct performance comparison with a parallel air aerated system. A sparged-turbine, gas-liquid contacting unit was employed in covered tanks of conventional design with discreet liquid and gas staging for cocurrent flow of each phase. This oxygenation system required very low power input for oxygen transfer and liquid mixing (0.08 - 0.140 HP/1000 gal. mixed-liquor) to routinely operate at dissolved oxygen concentrations of 8-10 mg/l while achieving > 90% utilization of feed oxygen gas. It was shown that the oxygenation system could operate with MLVSS concentrations of as high as 4500 mg/l achieving about 90% BOD removals at aeration detention times as low as 1.2 hours (raw flow + recycle flow) treating domestic waste of 220 mg/l BOD. Under other conditions and in direct comparison with air aeration, the oxygenation system consistently exhibited treatment performance superior to that achieved with air aeration even at uneconomically high aeration rates (3-5 cf air/gal. waste treated). Biomass from the oxygenation system is highly flocculent and readily settleable with desirable handling characteristics. Recycle and waste activated sludge suspended solids concentrations of about 3% were achieved.

Process advantages recognized with the oxygenation system indicate significant reductions in secondary waste treatment costs in comparison to conventional diffused air aerated systems. These advantages are principally reduced aeration tank volume requirements for equivalent treatment and a reduced production of waste activated sludge, a result of more efficient utilization of high purity oxygen than here-to-fore possible. As an example, the total cost of treatment, including preliminary treatment, with air at 6 hours nominal detention time and a 1.6 cf air/gal. aeration rate is estimated to be 15.1 cents, 13.0 cents and 11.0 cents/1000 gal. for plant sizes of 6, 30, and 100 MGD respectively while these costs are estimated to be reduced to 11.9 cents, 9.5 cents, and 7.8 cents/1000 gal. for respective 6, 30 and 100 MGD plant sizes with the use of an oxygenation treatment system capable of equivalent BOD removal at 2 hours nominal treatment detention time.

CONCLUSIONS AND RECOMMENDATIONS

A practical and economically attractive means for use of high purity oxygen gas in aeration of the conventional activated sludge process has been tested over an eight month period of continuous operation. This work was done using a temporarily installed oxygenation system in the existing aeration tanks at the Batavia, New York Municipal Pollution Control Plant. A sparged-turbine, gas-liquid contacting unit was employed for oxygen transfer. The aeration tanks, which were of conventional design, were covered and equipped with gas-liquid staging baffles to provide a multistage oxygenation system. This system functioned with co-current gas and liquid flow.

The test program was specifically planned to evaluate the following:

1. The feasibility and economics of high purity oxygen use in aeration of the conventional activated sludge waste treatment process.
2. The comparison of oxygenation and air aeration in terms of treatment performance.
3. The consistent operation of the oxygenation system at high mixed-liquor suspended solids (MLSS) levels (6000-8000 mg/l), high dissolved oxygen concentrations (8-10 mg/l) and high overall utilization of feed oxygen gas (>90%).
4. The operation and treatment performance of an oxygenation system at low detention times (approximately 1.2 hours based on raw wastewater + recycle sludge flow) and high organic loading conditions considered economically desirable but impractical with air aeration.
5. The relative economics of domestic waste treatment comparing the cost of oxygenation with conventional (diffused air) air aeration.

The work was conducted in three phases. In two phases, the performance of the oxygenation system was compared to a parallel air aeration system. In a third phase, the oxygenation system was used with one-fourth of the plant aeration tankage to treat the entire wastewater flow to the plant. A summary of results from each of the three phases of operation is shown in Table 1.

TABLE 1

SUMMARY COMPARISON OF AIR AND OXYGENATION SYSTEMS PERFORMANCE

	PHASE I OPERATION		PHASE II OPERATION	PHASE III OPERATION	
	AIR	OXYGEN	OXYGEN	AIR	OXYGEN
Wastewater Feed Rate (MGD)	1.97	1.91	2.53	1.29	1.44
Aeration Detention Time (Hrs.)*	3.6	3.6	1.2	2.6	2.0
Nominal Aeration Detention Time (Hrs.)**	4.0	4.1	1.5	3.0	2.8
MLSS Concentration (mg/l)	2440	3060	6980	3640	6190
MLVSS Concentration (mg/l)	1740	2210	4450	2580	4310
Recycle Sludge TSS (mg/l)	14,960	18,620	29,560	16,600	18,700
Volumetric Organic Loading lbs. BOD/Day/1000 ft^3	60.0	57.9	212.5	128.9	144.8
Mixed-Liquor Dissolved Oxygen Concentration (mg/l)	1.5	8.7	9.0	0.8	8.0
Ft^3 Air Utilized/Gal Sewage Treated	2.89	-	-	4.32	-
% Feed Oxygen Utilized		95.5	92.7	-	91.4
% BOD Removed	90	92	90	88	94
% COD Removed	76	80	71	79	84
% TSS Removed	93	96	89	94	97

* Raw Flow + Recycle Flow

** Raw Flow Only

The oxygenation system consistently exhibited superior treatment performance in direct comparison to the parallel air aeration system in spite of the fact that an impractically high air aeration rate was used to insure optimum performance. The power required by the oxygenation unit to accomplish high treatment efficiency was only 0.08 - 0.14 HP/1000 gal. of wastewater treated.

The biomass generated by the oxygenation system was highly flocculant and rapidly settleable, yielding sludge volume index values as low as an average of 36. The high recycle and waste solids total suspended solids concentration achieved (approximately 3-4% when clarifier operation was closely controlled) obviates the requirement for thickening of the waste activated sludge prior to further processing for disposal.

An extensive evaluation of the oxygenation system treatment costs in comparison to conventional air aeration reveals significant potential savings attributable to high purity oxygen use. These savings are due to several characteristics of the oxygenation system, but the major factors are reduced aeration tank volume requirements for equivalent treatment and a reduced production of waste activated sludge. As an example, the total cost of air aeration treatment at 6 hours nominal detention time (based on raw wastewater flow) and a 1.6 cf/gal aeration rate is estimated to be 15.1 cents, 13.0 cents, and 11.0 cents/1000 gal for plant sizes treating 6, 30, and 100 MGD of wastewater. For equivalent BOD removal with oxygen aeration at 2 hour nominal detention time (based on raw wastewater flow) these total treatment costs are estimated to be 11.9 cents, 9.5 cents, and 7.8 cents/1000 gal for the respective 6, 30, and 100 MGD plant sizes. Total treatment costs referred to here include investment and operating costs associated with preliminary and primary treatment as well as the investment and operating costs associated with secondary treatment by the activated sludge process and the disposal of primary and waste activated sludges.

The promising results of the work reported clearly indicate the advisability of further work to evaluate in detail certain characteristics of the oxygenation system not yet studied in sufficient depth. Specifically,

the existing temporary installation should be used to evaluate the following:

1. The biokinetics of the treatment process as revealed by careful control of sludge wasting, frequent analysis of soluble BOD and COD in feed wastewater, mixed-liquor filtrate and final clarifier effluent.

2. Flow control should be established for the system such that the measurements in (1) above can be made under more nearly steady state conditions.

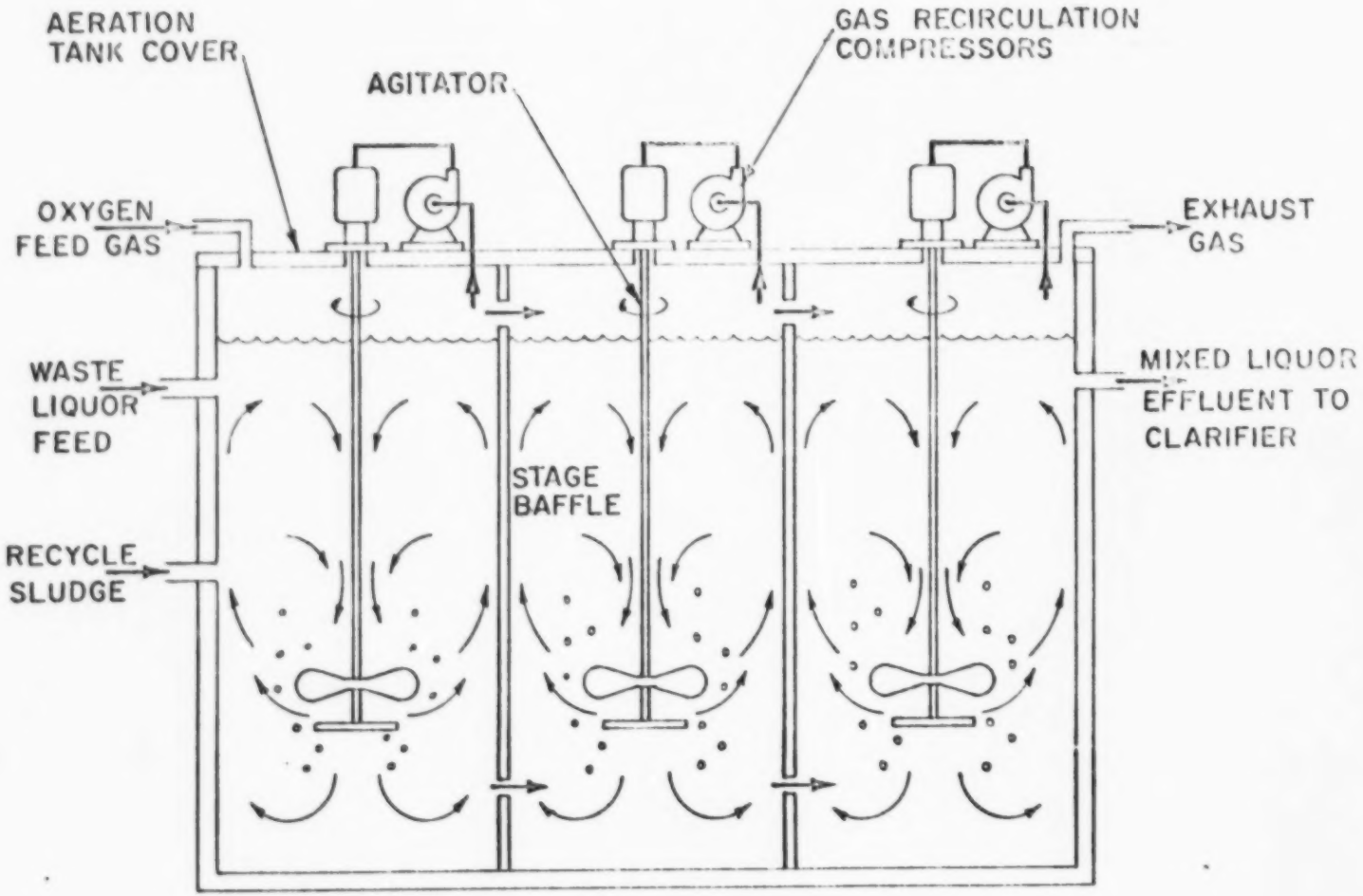
3. The handling properties of the waste activated sludge should be more fully elucidated by direct vacuum filtration tests.

4. System performance should be optimized by installation and operation of a flow proportion controlled activated sludge return system.

5. It would be desirable to evaluate oxygen aerated, aerobic digestion of waste activated sludge and to establish the handling characteristics of such stablized sludge by vacuum filtration tests.

FIGURE 1

SCHEMATIC DIAGRAM OF MULTI-STAGE
OXYGENATION SYSTEM



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E. K. Robinson	-	Contract Manager
J. G. Albertsson	-	Supervisor
N. P. Vahldieck	-	Senior Engineer
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D. V. Daly	-	Technician
J. R. Duemmer	-	Technician
S. E. Faruga	-	Technician
E. H. Kremer	-	Technician
P. F. Layer	-	Technician

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